COMBUSTION, SULPHUR NEUTRALISATION AND VANADIUM BALANCE TESTS WITH SYNCRUDE COKE AND FORT McMURRAY LIMESTONE IN A PILOT-SCALE ATMOSPHERIC RECIRCULATING FLUIDISED-BED REACTOR

FINAL REPORT

BY

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1. EXECUTIVE SUMMARY

A series of tests was conducted to evaluate the combustion of coke from the Syncrude tar sands plant in an atmospheric circulating fluidised-bed pilot-plant owned by Ahlstrom Corporation in Karhula, Finland. Key parameters investigated were: combustion efficiency; the effectiveness of limestone from Fort McMurray, Alberta, as a sulphur sorbent; sulphur capture; the fate and distribution of heavy metals contained in the coke. The test program was a follow-up to a previous exploratory test program carried out in 1981, in which Gotland (Sweden) limestone had been used as the sulphur sorbent.

The test program was carried out, under a contract with Supply and Services Canada, for CANMET, Energy, Mines and Resources Canada. The participants were Polymath Energy Consultants Ltd., Vancouver (prime contractor), Pyropower Corporation, San Diego, and the Ahlstrom Corporation, Karhula, Finland.

The test program utilised 50 tons of Syncrude fluid coke and 30 tons of Fort McMurray limestone in a series of 19 test runs. The coke had received no preparation prior to shipping. The limestone had been crushed and crated by the Alberta Research Council in Edmonton, Alberta. Combustion temperatures (nominal) in the tests were 850, 900 and 950°C. Runs were carried out with and without limestone injection to establish baseline SO_2 emissions and to obtain 90% sulphur capture. In 6 of the tests ash collected in the baghouse was reinjected into the combustor to ascertain the effect this had on sulphur capture and limestone utilisation. Since Syncrude coke contains about 6.5% sulphur, 0.20% vanadium and 0.06% nickel, effective capture of sulphur and heavy metals is critical if a commercial boiler is to be considered for the combustion of coke for steam and/or power generation.

The tests showed that the Fort McMurray limestone is fairly reactive, and 90% sulphur capture (range 88.6-91.4%) was achieved at Ca:S molar ratios ranging from 1.03:1 to 2.26:1. From the results, an optimum Ca:S molar ratio of 1.7:1 is recommended to guarantee 90% sulphur capture at 900°C in the combustor.

Baghouse ash recycling improved limestone utilisation and lowered the Ca:S molar ratio by about 10% for the sulphur capture obtained in the tests. The optimum baghouse ash recycle rate was established at about 67%.

Combustion efficiency was high (range 96.2-99.4% efficiency) at all temperatures within the test range. As the test proceeded, and the pilot-plant operators became more familiar with the fuel characteristics, the combustion efficiency improved. The last 9 test runs averaged 99% combustion efficiency (range 98.6-99.4%). Optimum combustion temperature was established as about 900°C.

NOx emissions could be minimised utilising a slightly lower combustion temperature (860-885°C) and careful control of the ratio of primary air to secondary air. Lowest NOx emissions of 50-70 ppm (0.032-0.045 mg/kJ) were obtained at primary/secondary air proportions below 60/40, i.e. when the primary air was less than 60% of the total combustion air. Under optimum conditions (900°C) NOx emissions were 220 ppm (0.128 mg/kJ).

Vanadium and nickel in the coke concentrated in the fly ash collected in the baghouse, with a significant amount also collecting in the boiler bottom ash. Vanadium and nickel in the particulates in the baghouse exhaust comprised less than 1% of the total input to the combustor. No attempt was made to measure gaseous vanadium or nickel compounds in the stack gases; this loss is expected to be minor.

Conclusions drawn from the test program are that Syncrude fluid coke can be burned successfully in a PYROFLOW circulating fluidised-bed boiler (CFBB) with high combustion efficiency, 90% sulphur capture, low NOx emissions and low vanadium and nickel escape to the atmosphere. A commercial boiler can be designed to operate reliably on this fuel and should give no unusual operating problems.

RAPPORT ADMINISTRATIF

Une série d'essais a été effectuée afin d'étudier la combustion du coke provenant de l'usine de sable bitumineux de Syncrude; ces essais ont été réalisés dans l'usine pilote utilisant une chaudière à lit fluidisé avec circulation à pression atmosphérique de l'Ahlström Corporation, Karhula, Finlande. Les paramètres clé étudiés étaient les suivants: capacité de combustion; efficacité de la pierre à chaux provenant de Fort McMurray comme absorbant du soufre ; capture du soufre ; transformation et distribution des métaux lourds contenus dans le coke. Ce programme d'essais était la continuation d'un programme expérimental qui avait eu lieu en 1981 et pour lequel la pierre à chaux de Gotland (Suède) avait été utilisée comme absorbant du soufre.

Ce programme d'essais fut réalisé dans le cadre d'un contrat avec Approvisionnements et Services Canada, pour CANMET, Energie, Mines et Ressources Canada. Polymath Energy Consultants Ltd., Vancouver, (l'entrepreneur principal), Pyropower Corporation, San Diego, et Ahlström Corporation, Karhula, Finlande, participaient à ce programme.

Durant ce programme, 50 tonnes de coke Syncrude et 30 tonnes de pierre à chaux Fort McMurray furent utilisées pour une série de 19 essais. Les températures de combustion variaient entre 850, 900 et 950°C (seulement). Les essais furent effectués avec et sans injection de pierre à chaux afin d'établir le taux de base d'émanation de SO_2 et d'obtenir une capture de soufre de 90%. Lors de 6 de ces essais, les cendres recueillies dans la station de réception furent ré-injectées dans la centrale de combustion afin de pouvoir établir l'effet de cette opération sur la capture du soufre et l'exploitation de la pierre à chaux. Etant donné que le coke Syncrude contient environ 6.5% de soufre, 0.20% de vanadium et 0.06% de nickel, une efficace capture du soufre et des métaux lourds est cruciale si on envisage d'utiliser une chaudière commerciale pour la combustion du coke afin de produire de la vapeur d'eau et/ou de l'énergie.

Les essais ont prouvé que la pierre à chaux de Fort McMurray était assez réactive et qu'une capture du soufre de 90% (variant entre 88.6% et 91.4%) avait été obtenue avec des rapports molaires Ca:S allant de 1.03:1 à 2.26:1. D'après ces résultants, un rapport molaire Ca:S de 1.7:1 est recommandé afin d'assurer une capture du soufre de 90% à 900°C dans la chambre de combustion.

Le recyclage des cendres recueillies dans la station de réception a amélioré l'exploitation de la pierre à chaux et réduit le rapport molaire Ca:S de la capture du soufre (variant entre 88.6% et 91.4%). obtenue durant les essais, d'environ 10%. Le taux optimum de recyclage des cendres recueillies dans la station de réception a été établi à environ 67%.

Pour toutes les températures utilisées dans ce programme d'essais, le rendement de combustion fut élevé (variant entre 96.2% et 99.4%). Au fur et à mesure que les essais se déroulaient, les opérateurs de l'usine pilote se familiarisaient avec le combustible et le rendement de combustion s'est, de ce fait, nettement amélioré.

Les derniers 9 essais ont donné une moyenne de rendement de combustion d'environ 99% (variant de 98.6% à 99.4%). La température de combustion optimum a été établie à environ 900°C.

Les émanations de NOx ont pu être minimisées en utilisant une température de combustion légèrement plus basse (860-885°C) et en exerçant un contrôle attentif sur le taux d'air primaire et le taux d'air secondaire. Les émanations les plus faibles de NOx, 50-70 ppm (0.032-0.045 mg/kJ) ont été obtenues avec une proportion d'air primaire et secondaire inférieure à 60/40, c'est à dire que l'air primaire représentait moins de 60% de l'air total de combustion. Dans les conditions optimum (900°C), les émanations de NOx étaient de 220 ppm (0.128 mg/kJ).

Le vanadium et le nickel présents dans le coke se concentraient dans les cendres volantes recueillies dans la station de réception, un taux important fut également recueilli au fond de la chaudière. Le vanadium et le nickel présents dans les particules recueillies au niveau de l'évacuation de la station de réception représentait moins de 10% de l'apport total au niveau de la centrale de combustion. Aucune tentative ne fut faite pour mesurer les composés de vanadium ou de nickel gazeux au niveau des gaz d'évacuation; on pense que cette perte est mineure.

Les conclusions que l'on peut tirer de ce programme d'essais sont les suivantes: le coke Syncrude peut être consummé avec des résultats satisfaisants dans une chaudière PYROFLOW à lit fluidisé avec circulation à pression atmosphérique (CFBB); on obtient ainsi un haut rendement de combustion, une capture du soufre de 90%, de faibles émanations de NOx et un échappement minimal de vanadium et de nickel dans l'atmosphère. Une chaudière commerciale peut être conçue afin de fonctionner de façon fiable avec ce combustible, sans poser de problèmes particuliers.

2. INTRODUCTION

Syncrude Canada Ltd. operates a tar sands extraction plant at Fort McMurray, Alberta. Tar sands are mined and treated in the plant to produce heavy, synthetic crude oil. The Syncrude plant includes fluid coking units which recover lighter hydrocarbon fractions as saleable products. A considerable amount of carbon is discharged from the cokers as a solid coke waste product (2,500 Mt/day).

The coke has a high calorific value but is difficult to utilise by conventional combustion technology because of its high sulphur content (6-7%), low volatile content (6%) and some carbon forms which are less reactive. The coke contains high concentrations of heavy metals, particularly vanadium (0.17-0.20%) and nickel (0.06%), which can cause severe corrosion problems in conventional boilers.

In 1981, Syncrude coke was combusted at the Hans Ahlstrom laboratory, Finland. The combustion efficiencies varied between 96% and 99%, and 90% sulphur capture was obtained with a Ca:S molar ratio of about 2:1. Gotland limestone (from Sweden) was used as the sulphur sorbent.

In 1985 CANMET decided to initiate a further series of combustion tests on the coke to obtain more data than had been possible during the earlier (1981) test program. Data from the new tests are to be useful for optimising the design of a commercial-scale CFBB unit. In particular, limestone from Fort McMurray was to be evaluated as a sulphur sorbent, and the mass balance and distribution of vanadium and nickel were to be determined accurately.

A contract was awarded to Polymath Energy Consultants Ltd. to undertake a combustion test program in the PYROFLOW 1.5 MW circulating fluidised-bed pilot-plant of the Ahlstrom Corporation at their Hans Ahlstrom laboratory, Karhula, Finland. The project was undertaken cooperatively by Polymath Energy Consultants Ltd., Ahlstrom, and the Pyropower Corporation of San Diego, California.

For the contract, 50 tons of Syncrude coke and 30 tons of Fort McMurray limestone were shipped to Finland. The tests were run between 5 December 1985 and 17 January 1986. The coke was burned as received. The limestone was pulverised to -0.50 mm and dried before use as the sulphur sorbent.

The test conditions, results and conclusions drawn from the test program are discussed in this report.

3. TEST OBJECTIVES

The main objective of the test was to study the combustion of Syncrude coke in a circulating fluidised-bed combustor (CFBC). The following particular objectives, as specified in the contract, were set for the test program.

A. to establish base-level emissions of SO_2 when burning Syncrude coke without limestone;

B. to establish the effect of bed temperature and superficial gas velocity on combustion efficiency and sulphur retention when using limestone as the sulphur sorbent;

C. to establish the effect of baghouse ash recycle on combustion efficiency and sulphur retention;

D. to establish the combination of bed temperature, superficial fluidising velocity, primary to secondary air flow ratio, and baghouse ash recirculation rate which produce optimum combustion efficiency and limestone utilisation; and

E. to determine the fate of heavy metals in the coke, particularly vanadium and nickel.

4. DESCRIPTION OF TEST FACILITIES

All combustion tests were conducted in the atmospheric circulating fluidised-bed PYROFLOW pilot-plant of the Ahlstrom Corporation in Karhula, Finland. This unit has a thermal rating of approximately 1.5 MW (1.5 MJ/s). A schematic diagram of the pilot-plant is shown on Figure 1. Principal components of the pilot-plant, and a description of their function, are presented below.

4.1 Combustor (reactor)

This is a refractory-lined cylinder, 600 mm internal diameter and about 8.5 m high. At its base a windbox supplies high pressure primary air to the combustor through a bed support plate. Secondary combustion air can be supplied at five levels in the combustor freeboard space through two diametrically opposed air nozzles at each level. High pressure air (up to 250 mB) is supplied to the combustor from two centrifugal fans running in series.

The combustor is equipped with oil and gas burners for start-up. Solid fuel feed is started when the combustor temperature reaches about 600°C. Solid fuel is fed from a weighed feed hopper via a screw conveyor and rotary valve into the loop seal ash return flow to the combustor. Limestone, for SO_2 absorption, may be fed from a hopper via a screw conveyor and rotary valve into the loop seal adjacent to the fuel feed point.

The high air pressure and air flow rate/unit area causes the mass transport of fine bed material from the dense bed region into the combustor freeboard space, thence from the combustor into an external cyclone. The turbulent suspension of bed material in the freeboard space gives a high degree of radiant heat production in the upper part of the combustor. Heat is removed from the combustion chamber, and temperature is partially controlled, by the use of water-cooled bayonet coolers inserted into the top of the combustion chamber through gas-tight seals/clamps.

In commercial PYROFLOW units, heat removal from the combustor is via conventional membrane water walls with water tubes aligned parallel to the gas flow. Since the Karhula unit is only a pilot-plant, bayonet coolers are used to remove heat from the radiant zone. No in-bed tubes are used for heat removal.

Bed ash is removed via an air cooled ash cooler, which returns ash fines to the combustor and discharges cooled ash to the ash disposal system. The ash cooler may also be water cooled if required.

Temperature and pressure measurements are made at numerous points within the combustor. These include the dense bed region just above the support plate, and at various levels in the combustor freeboard space and in the gas exit duct.

Fuel feed rate is measured by loss of weight of the feed hopper, with a cross-check by screw feeder speed after it has been calibrated for each fuel. Limestone feed rate is measured by screw feeder speed and by an impact device that measures the force exerted on a hinged plate by the falling stream of limestone.

Ash may be recycled to the combustor from the baghouse via a pneumatic delivery system. Ash recycle rate can be varied from the control room.

4.2 Cyclone

A refractory-lined cyclone is connected in series with the combustor. Gases and bed materials leaving the combustor are separated in the cyclone, with the solid materials falling to the base of the cyclone for subsequent reinjection into the combustor.

The cyclone operates at a high temperature (up to reactor temperature) and allows additional residence time for carbon burn-up of elutriated fuel particles. Evidence of some additional combustion occurring in the cyclone is provided by a slight temperature increase from the combustor exit to the cyclone exit.

The cyclone is about the same diameter as the combustor (600 mm) but is only about 3 m high. The top of the cyclone is level with the top of the combustor.

At the base of the cyclone a vertical tube delivers solid material to a non-mechanical loop seal. This seal transfers solids from the cyclone back into the combustor. The loop seal is designed to allow the ash to flow back to the combustor, while blocking the flow of combustion gases and bed material into the base of the cyclone.

No heat exchange surfaces are installed in the cyclone. Its functions are solely as a secondary burn-out chamber for unburned carbon and as a solids separator for circulating bed material. Temperature and pressure are measured in the cyclone.

4.3 Boiler

Gases exiting the cyclone, at a temperature of 850-950°C, enter a gas/water heat exchanger (boiler) with water tubes aligned normal to the gas flow. Water is circulated through the tubes to cool the gas to about 200°C. Water flows and temperatures are measured and recorded. The gas flow is directed via internal baffles to make several passes over the tubes before leaving the boiler.

The boiler is equipped with air-operated soot blowers. Tube deposits and coarser suspended particles settle into the base of the boiler casing, from where they are discharged into a weighed ash hopper via a valve. The ash hopper discharges ash to an outside ash silo by a pneumatic transport line.

4.4 Baghouse

A two-compartment baghouse with air pulse cleaning removes fine particulates to an acceptable level for stack discharge. The bag elements are folded cartridges made from a Nomex-type fabric. Normal operating temperature is 200-220°C, with a maximum temperature of 300°C.

Ash falls into a hopper in the base of the baghouse and is removed through air-operated ball valves. Ash may be directed into a weighed bin below the baghouse or into a screw feeder and rotary valve for pneumatic transport back to the combustor. Ash recycle rate is determined by varying the open/close frequency and timing of the air-operated ball valves from the control room. Ash from the weighed ash bin is delivered to a pneumatic transport line for disposal to an outside ash silo. A valved sample line permits the collection of ash samples for analysis.

During start-up, when gas temperature is low and near the dew point, the baghouse is by-passed. During these occasions, and also during emergencies or problems with the baghouse, dust removal is accomplished using a venturi scrubber.

4.5 Venturi scrubber

All gas flow enters a vertical venturi scrubber prior to the induced draft fan and stack. The venturi removers fine particulates escaping the baghouse, and is also used for flue gas cleaning during start-up, when the baghouse is by-passed until the flue gas temperature is high enough to prevent condensation.

The venturi is a low gas-pressure drop scrubber (8-10 mB) that utilises a high pressure water jet to entrain the gas. Solids accumulate in a settling/collection chamber beneath the venturi.

4.6 Control room

Pilot-plant data logging and readout is centralised in a control room adjacent to the pilot-plant. Process data measurements are made by an ALCONT automatic process control system. Pilot-plant status is presented on control readout terminal (CRT) displays featuring an overall flow schematic for the system as well as several sub-loops that can be called up on demand. All relevant temperature, pressure, flow, hopper weight, etc. data are shown on the CRT and are also recorded on multi-pen charts above the CRT displays. A 30 minute print-out of time averaged values is also available for data analysis.

All pilot-plant functions can be monitored and controlled from the computer terminals in the control room.

4.7 Other features

Process monitoring

Points at which temperatures, pressures, flows, weights, etc. are measured and displayed on the control room CRTs are illustrated on Figures 2 to 4. All functions can be controlled from the central control room.

Gas monitoring and analysis

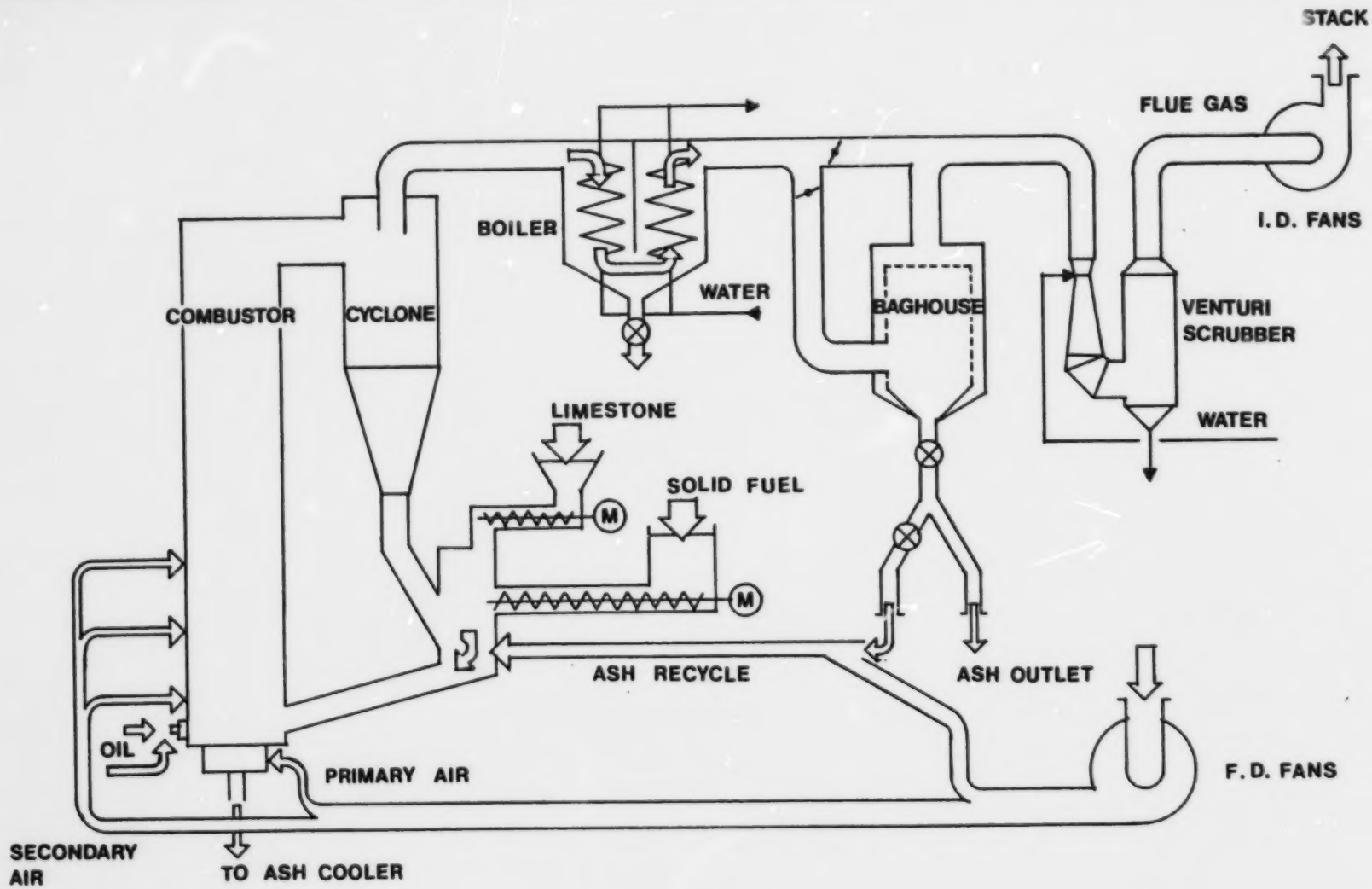
Flue gas monitoring locations and equipment are shown on Figure 5. All instrument readings are recorded on strip-chart recorders, with excess O_2, CO, SO_2, and NOx also being displayed on the control room CRTs. Note that particulate samples are collected by an isokinetic sampler located in the baghouse exhaust duct; when the baghouse is by-passed high particulate levels are measured.

Instrumentation includes:

- Westinghouse and Taylor excess O_2 analysers
- Teco CO analyser
- Teco, Uras and Monitor Labs SO_2 analysers
- Beckman and Monitor Labs NOx analysers
- J.U.M. total hydrocarbon analyser
- Severn Science SO_3 analyser
- Isokinetic particulate sampler (non-recording manual device)

Solids sampling

Sampling points are shown on Figure 6. Samples are collected directly from each stream into a steel container.

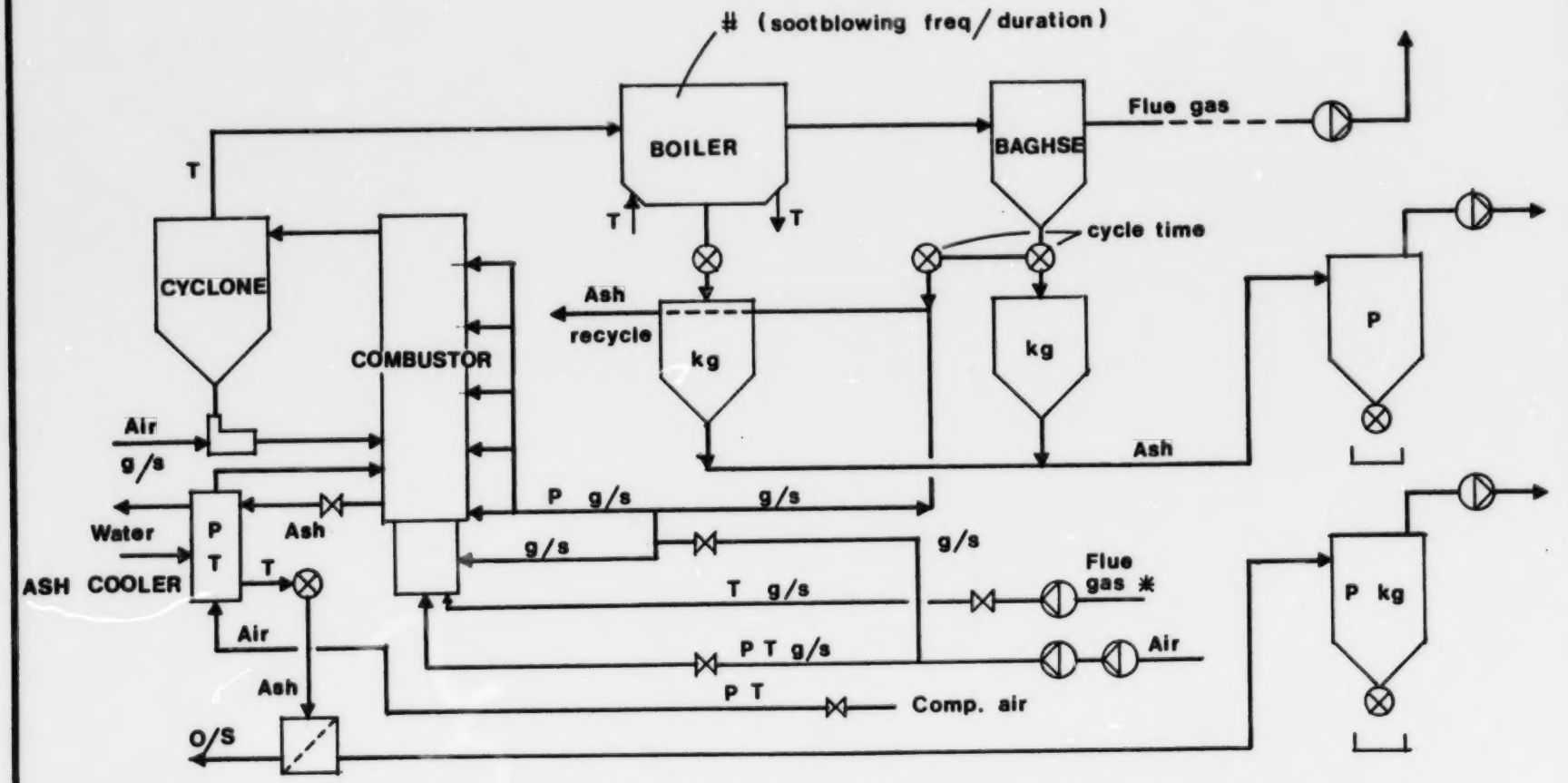


⊗ – Rotary valve

Ahlström PYROFLOW 600 mm pilot-plant schematic

Polymath Energy Consultants Ltd.

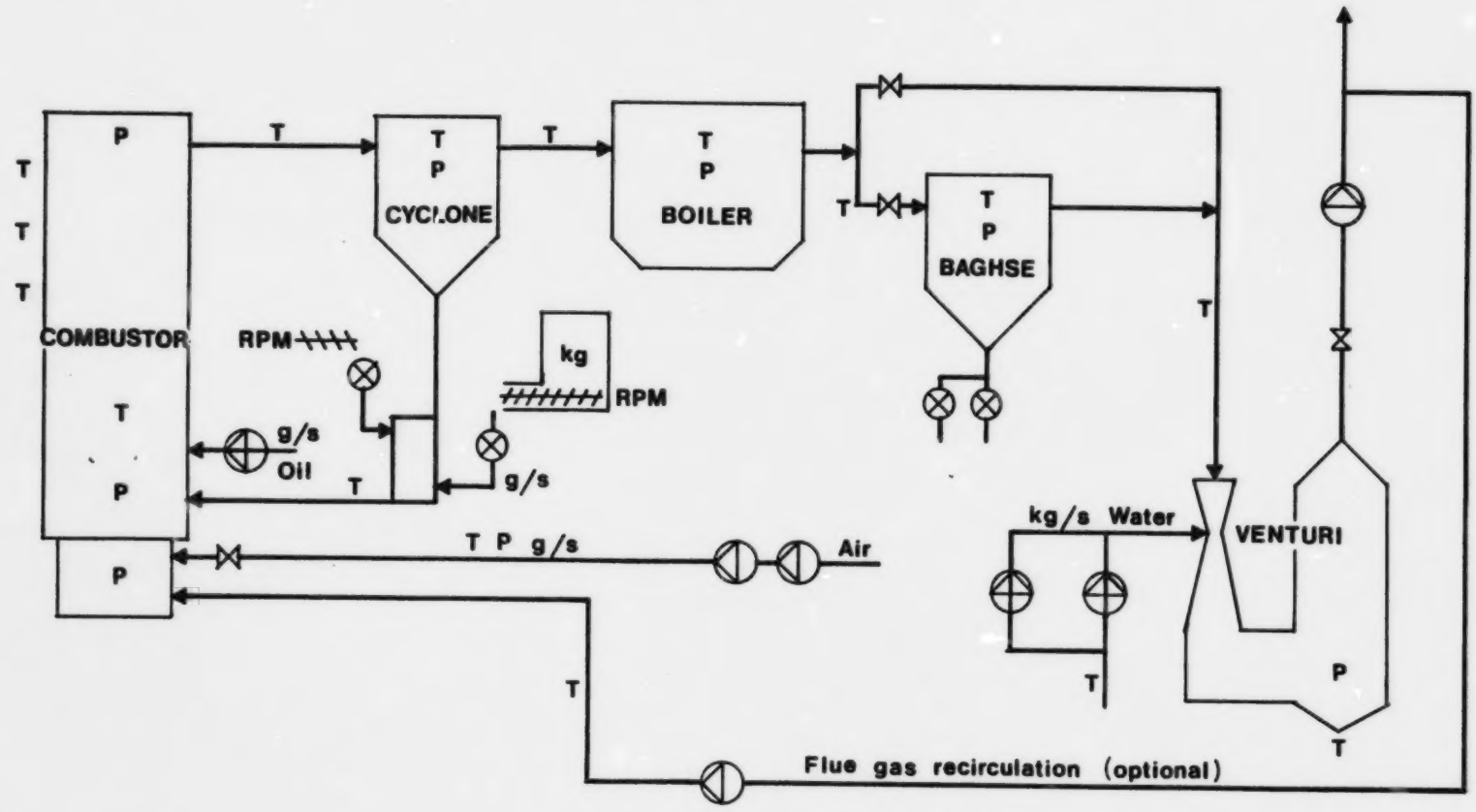
Figure 1



Air, gas and ash flow and measurement schematic

Polymath Energy Consultants Ltd.

Figure 2



⊗ – Rotary valve

T, P – Temperature, Pressure

g/s , kg/s – Flow measurement

– Fan

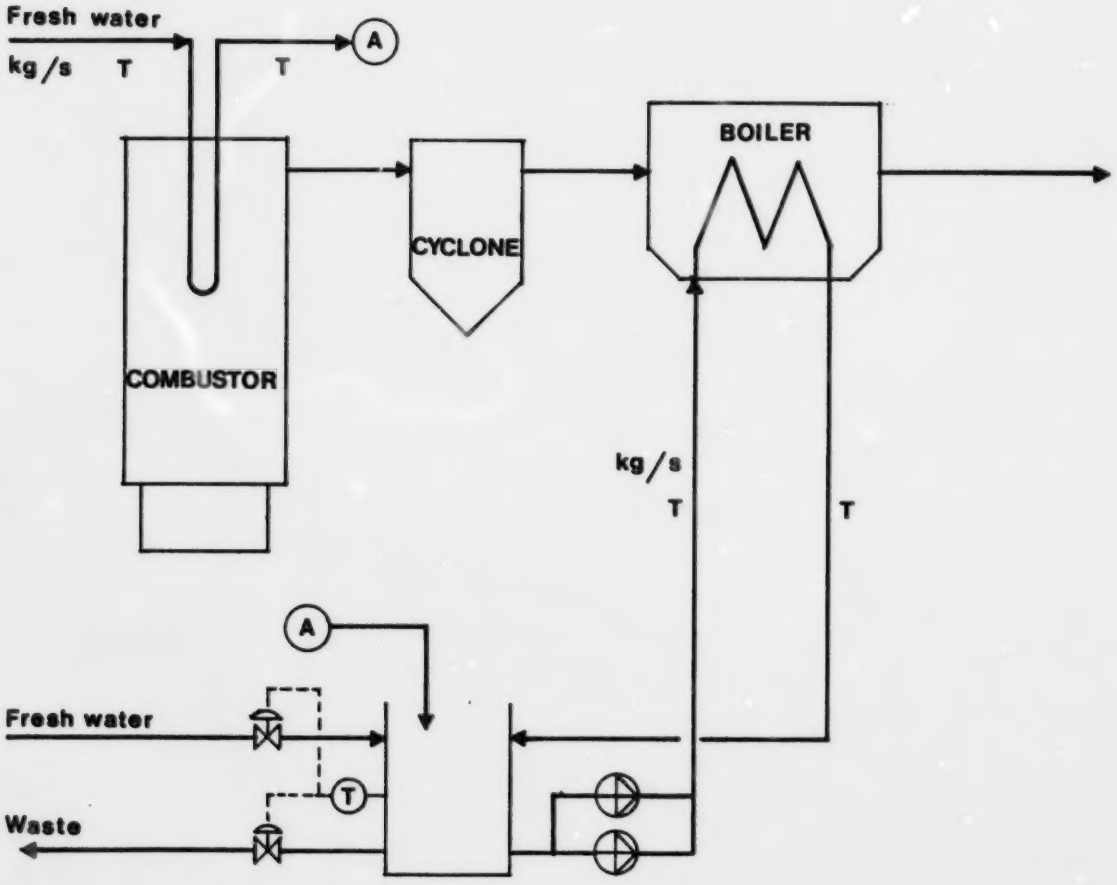
– Pump

RPM – Feeder speed

Temperature, pressure, flow measurement schematic

Polymath Energy Consultants Ltd.

Figure 3



– Control valve

T – Temperature

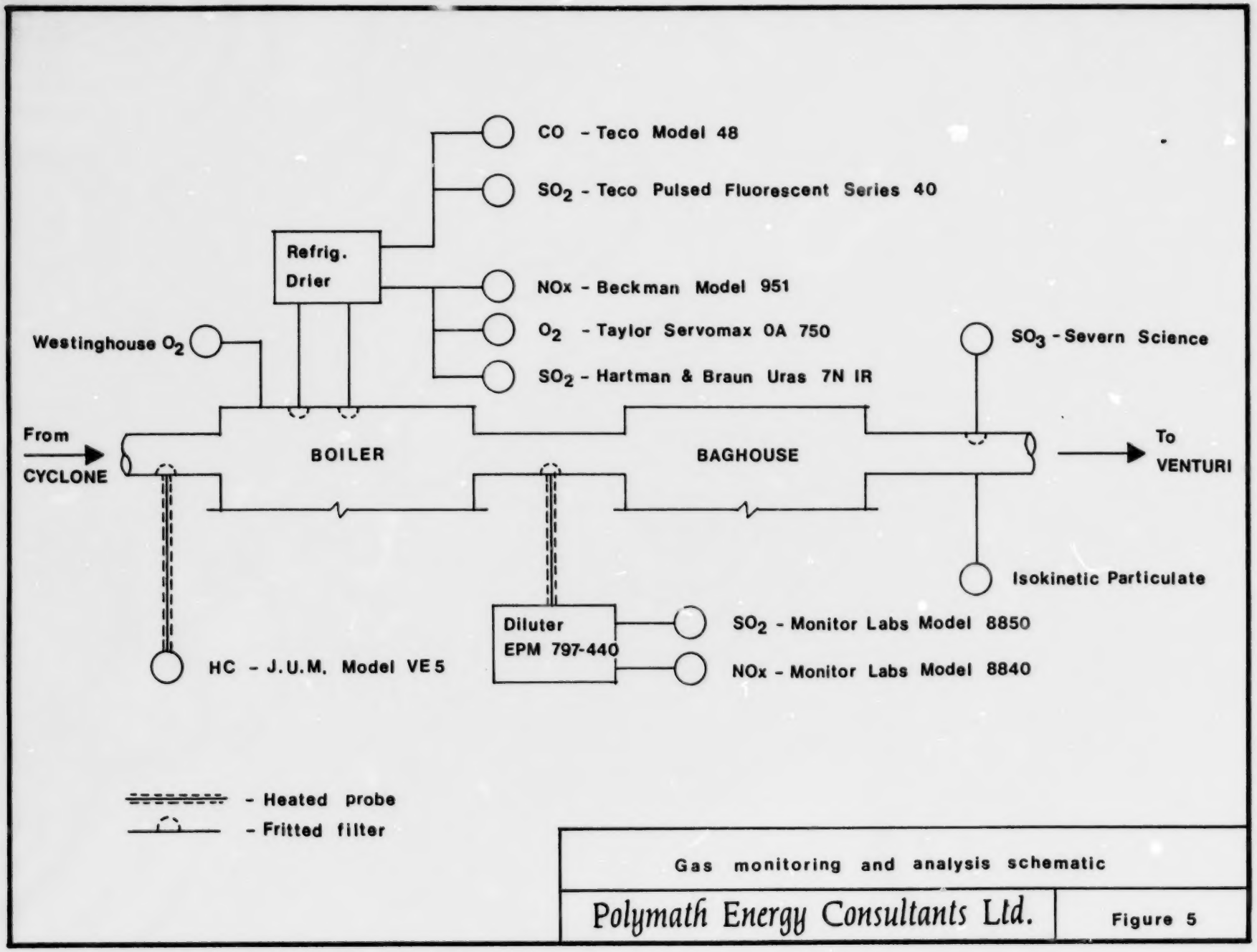
kg/s – Flow measurement

– Pump

Cooling water flow schematic

Polymath Energy Consultants Ltd.

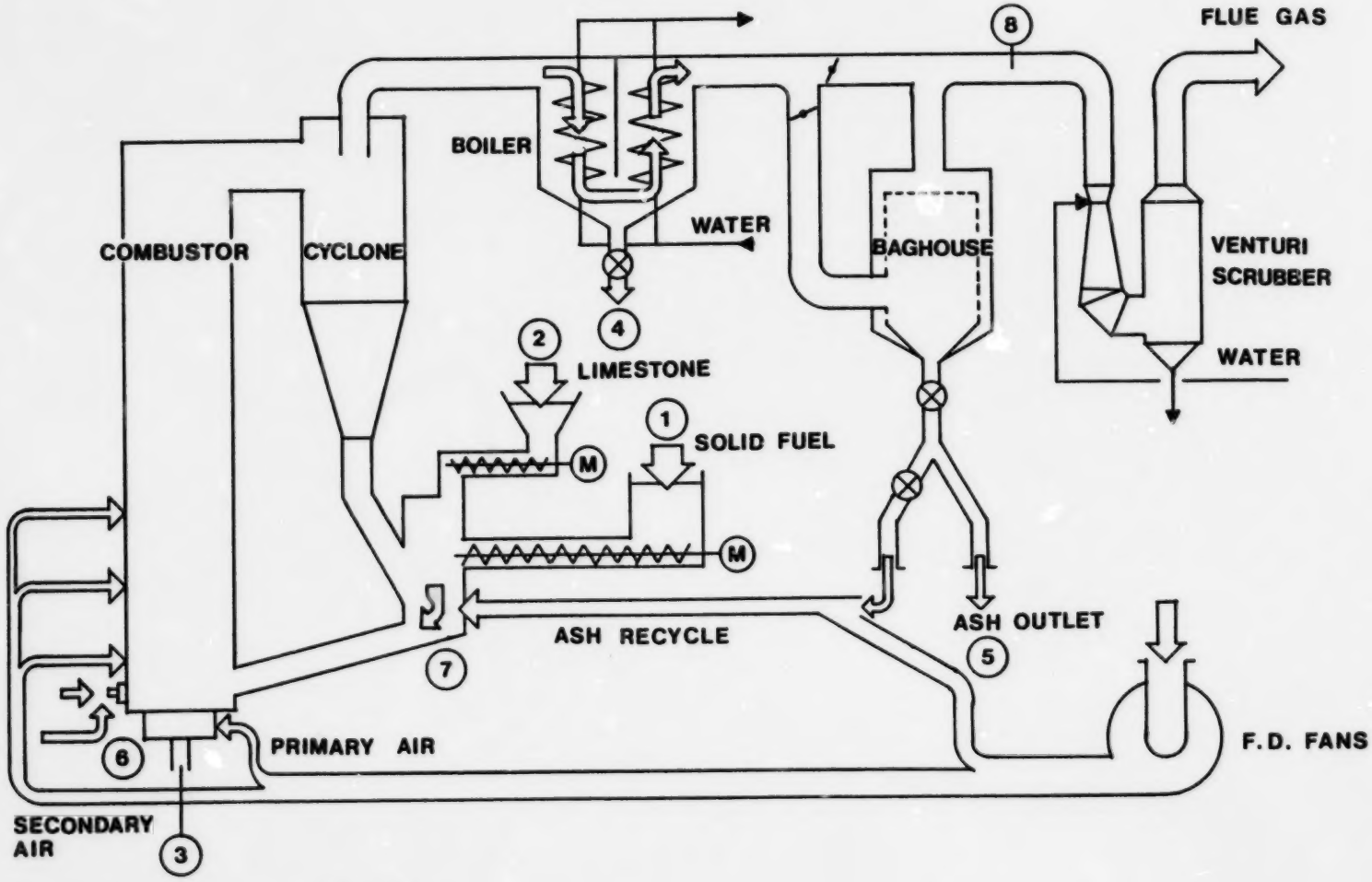
Figure 4



Gas monitoring and analysis schematic

Polymath Energy Consultants Ltd.

Figure 5



1 - Coke feed
2 - Limestone
3 - Bottom ash
4 - Boiler ash
5 - Baghouse ash
6 - Bed material
7 - Circulating material
8 - Particulate sampler

Ahlström PYROFLOW pilot-plant sampling points

Polymath Energy Consultants Ltd.

Figure 6

5. TEST PROCEDURE

5.1 Test plan

The planned test program consisted of 17 separate test runs. Run No. 0 was a scoping test to establish reliable operating values for the system. Runs 1-3 were baseline tests without limestone addition using different combustion temperatures. In the rest of the test runs (4-16), Fort McMurray limestone was used as a sorbent in order to achieve 90% sulphur capture. Temperature, baghouse ash recirculation and firing rate are variables. The main test parameters are given in Table 1.

Table 1 Main test parameters

Run No.	Bed temp. target °C	Load cond.	Limestone addition	Baghouse ash recycle	Sulphur ret. target %
0	900	Nominal	No	No	-
1	850	Nominal	No	No	-
2	900	Nominal	No	No	-
3	950	Nominal	No	No	-
4	950	Nominal	Yes	No	90
5	900	Nominal	Yes	No	90
6	850	Nominal	Yes	No	90
7	850	Nominal	Yes	Yes	90
8	900	Nominal	Yes	Yes	90
9	950	Nominal	Yes	Yes	90
10	950	Maximum	Yes	No	90
11	950	Maximum	Yes	Yes	90
12	900	Intermed.	Yes	Yes	90
13	900	Intermed.	Yes	No	90
14	850	Minimum	Yes	No	90
15	850	Minimum	Yes	Yes	90
16	Optimum	Nominal	Yes	Optimum	90

The most important goals in all of the test runs were to measure the Ca:S molar ratio and combustion efficiency, and to determine the vanadium and nickel balances.

As will be described in this report, the planned test program was expanded to 19 test runs because of temporary mechanical problems with the fly ash recirculation equipment.

5.2 Trials in chronological order

The Syncrude coke combustion tests were started on 26 November, 1985, by calibrating the instruments and the analysers and heating up the pilot-plant.

During the initial test, i.e. the scoping test run No. 0, it was noticed that the coke contained stones that disturbed the feeding and damaged the rotary feed valve. The feeding screw had to be changed and the rotary feeder repaired.

A screen was installed before the coke feeding hopper to separate stones from the coke. The test program was continued on 5 December, 1985. Tests 0 to 3 inclusive were without limestone addition, to establish baseline SO_2 emissions, and tests 4 to 16 inclusive were with the addition of limestone for sulphur capture. Note that the test runs were not consecutive according to run number. The actual order of the tests, as performed, was: 0, 1, 2, 3, 4, 5, 8A, 7A, 6, 7B, 8B, 9, 14, 15, 12, 13, 11, 10, 16.

Test run No. 0
December 5, 1985. 1500-2100 h*.

Run 0 was a scoping test to establish reliable operating values. The mean reactor (combustor) temperature was 896°C, and average excess O_2 was 3.5% (dry basis)**. Superficial fluidising velocity (SFV) was 4.6 m/s.

Test run No. 1
December 10, 1985. 0500-0800 h.

Run 1 was a short test at low temperature. The mean reactor temperature was 823°C. Gas analysis values were: O_2 = 3.6%; SO_2 = 4,300 ppm; CO = 650 ppm; NOx = 100 ppm. SFV was 4.7 m/s.

Test run No. 2
December 11, 1985. 0300-1400 h.

Run 2 was an 11 hour test at a mean reactor temperature of 890°C. O_2 averaged 3.2%; SO_2 = 4,800 ppm; CO = 108 ppm; NOx = 150 ppm SFV was 4.7 m/s.

Test run No. 3
December 12, 1985. 0000-1900 h.

Run 3 was the third and final baseline test, with a mean reactor temperature of 940°C. O_2 averaged 3.8%; SO_2 = 5,000 ppm; CO = 80 ppm; NOx = 520 ppm. SFV was 5.9 m/s.

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* All times cover the steady-state portion of each test, after the reactor had been brought to equilibrium.
** All gas analyses expressed on dry basis.

Test run No. 4
December 17, 1986. 0300-1200 h.

This was the first test run using limestone for sulphur capture. Fort McMurray limestone (from Fort McMurray, Alberta; used in tests 4-16) was fed at 20.1 g/s, corresponding to a Ca:S molar ratio of 2.04. The mean reactor temperature was 937°C. SO_2 emissions dropped to 480 ppm (89.9% sulphur capture). O_2 averaged 4.7%; CO = 50 ppm; NOx = 310 ppm. SFV was 6.1 m/s.

Test run No. 5
December 18, 1985. 1430-2300 h.

The mean reactor temperature was lowered to 881°C. Limestone was fed at 15.4 g/s, giving a Ca:S molar ratio of 1.54. SO_2 emission averaged 500 ppm (90.2% sulphur capture). O_2 = 3.2%; CO = 70 ppm; NOx = 130 ppm. SFV was 5.6 m/s.

Baghouse ash recirculation into the reactor was planned for test runs 7 and 8. Technical problems with the recirculation system resulted in two runs (7A, 8A) being performed without ash recirculation. After the recirculation system was repaired, test runs 7 and 8 were renewed (7B, 8B) with baghouse ash recirculation into the reactor.

Test run No. 8A
December 19, 1985. 0650-1445 h.

The mean reactor temperature was 884°C. Limestone was fed at 16.9 g/s, giving a Ca:S molar ratio of 1.71. SO_2 emission averaged 490 ppm (89.6% sulphur capture). O_2 = 4.5%; CO = 40 ppm; NOx = 120 ppm. SFV was 5.3 m/s.

Test run No. 7A
December 20, 1985. 1000-1930 h.

The mean reactor temperature was lowered to 833°C. Limestone was fed at 20.4 g/s, giving a Ca:S molar ratio of 2.11. SO_2 emission averaged 460 ppm (91.0% sulphur capture). O_2 = 3.5%; CO = 120 ppm; NOx = 160 ppm. SFV was 4.9 m/s.

After test run No. 7A the pilot-plant was shut-down over the Christmas and New Year holiday period. The test program resumed on 7 January, 1986.

Test run No. 6
January 7, 1986. 0530-1345 h.

Mean reactor temperature was 854°C. Limestone was fed at 21.3 g/s, giving a Ca:S molar ratio of 2.26. SO_2 emission averaged 480 ppm (90.7% sulphur capture). O_2 = 3.4%; CO = 90 ppm; NOx = 200 ppm. SFV was 5.3 m/s.

Test run No. 7B
January 9-10, 1986. 2030-0300 h.

This was the first test with baghouse ash recirculation to the reactor; 80% of the baghouse ash was recycled to the reactor. Mean reactor temperature was 842°C. Limestone was fed at 14.7 g/s, giving a Ca:S molar ratio of 1.64. SO_2 emission averaged 470 ppm (90.0% sulphur capture). O_2 = 5.0%; CO = 30 ppm; NOx = 175 ppm. SFV was 5.3 m/s.

Test run No.8B
January 10, 1986. 0630-1250 h.

80% of baghouse ash was recirculated into the reactor. Mean reactor temperature was 896°C. Limestone was fed at 15.0 g/s, giving a Ca:S molar ratio of 1.65. SO_2 emission averaged 480 ppm (90.1% sulphur capture). O_2 = 4.3%; CO = 45 ppm; NOx = 185 ppm. SFV was 5.6 m/s.

Test run No. 9
January 10, 1986. 1430-2100 h.

80% of baghouse ash was recirculated to the reactor. Mean reactor temperature was raised to 938°C. Limestone was fed at 17.6 g/s, giving a Ca:S molar ratio of 1.72. SO_2 emission averaged 490 ppm (90.6% sulphur capture). O_2 = 3.2%; CO = 40 ppm; NOx = 245 ppm. SFV was 5.8 m/s.

Test run No. 14
January 14, 1986. 0900-1305 h.

This was a minimum load test without baghouse ash recycle. Mean reactor temperature was 862°C. Limestone was fed at 6.4 g/s, giving a Ca:S molar ratio of 1.21. SO_2 emission averaged 580 ppm (88.6% sulphur capture). O_2 = 3.6%; CO = 45 ppm; NOx = 54 ppm. SFV was 3.1 m/s.

Test run No. 15
January 14, 1986. 1330-1800 h.

Run 15 was a minimum load test with 80% baghouse ash recycled to the reactor. Mean reactor temperature was 875°C. Limestone was fed at 5.0 g/s, giving a Ca:S molar ratio of 1.03. SO_2 emission averaged 520 ppm (90.0% sulphur capture). O_2 = 3.2%; CO = 50 ppm; NOx = 97 ppm. SFV was 2.8 m/s.

Test run No. 12
January 14-15, 1986. 2015-0030 h.

This was an intermediate load test with 80% baghouse ash recirculation to the reactor. Mean reactor temperature was 883°C. Limestone was fed at 9.9 g/s, giving a Ca:S molar ratio of 1.34. SO_2 emission averaged 490 ppm (89.9% sulphur capture). O_2 = 4.8%; CO = 20 ppm; NOx = 70 ppm. SFV was 4.0 m/s.

Test run No. 13
January 15, 1986. 0100-0750 h.

This was an intermediate load test without baghouse ash recirculation to the reactor. Mean reactor temperature was 882°C. Limestone was fed at 9.0 g/s, giving a Ca:S molar ratio of 1.41. SO_2 emission averaged 440 ppm (90.8% sulphur capture). O_2 = 5.0%; CO = 10 ppm; NOx = 65 ppm. SFV was 3.8 m/s.

Test run No. 11
January 15, 1986. 0900-1200 h.

Run 11 was a maximum load test with 80% baghouse ash recirculation to the reactor. Mean reactor temperature was 960°C. Limestone was fed at 20.2 g/s, giving a Ca:S molar ratio of 1.89. SO_2 emission averaged 430 ppm (91.4% sulphur capture). O_2 = 4.0%; CO = 5 ppm; NOx = 450 ppm. SFV was 6.9 m/s.

Test run No. 10
January 15, 1986. 1530-1900 h.

This was a maximum load test without baghouse ash recycle. Mean reactor temperature was 947°C. Limestone was fed at 19.3 g/s, giving a Ca:S molar ratio of 1.73. SO_2 emission averaged 470 ppm (90.9% sulphur capture). O_2 = 3.5%; CO = 20 ppm; NOx = 240 ppm. SFV was 6.3 m/s.

Test run No. 16
January 16-17, 1986. 0900-1000 h.

Run 16 was a long test (25 hours) under optimal conditions. Mean reactor temperature for this test was 900°C. 67% of baghouse ash was recirculated into the reactor. Limestone was fed at 14.7 g/s, giving a Ca:S molar ratio of 1.53. SO_2 emission averaged 480 ppm (90.5% sulphur capture). O_2 = 4.0%; CO = 10 ppm; NOx = 220 ppm. SFV was 5.5 m/s.

After run 16 the pilot-plant was shut down as all the Syncrude coke had been burned in the tests.

6. RESULTS

6.1 Coke characteristics

The analysis of blended grab samples of the coke as received (taken from the crates by thief sampler) by the Technical Research Centre of Finland is given in Table 2.

Table 2. Coke analysis

Proximate analysis:	
- moisture	3.41%
- volatiles in dry solids	5.6 %
- ash in dry solids	9.5 %
- fixed carbon in dry solids	84.9 %
Ultimate analysis (wt-% in dry solids)	
- C	79.6 %
- H	1.3 %
- N	1.4 %
- S	6.6 %
- ash	9.5 %
- O (by difference)	1.6 %
Higher heating value	28.56 MJ/kg (dry basis)
ASTM ash fusion temperatures:	
- oxidising conditions	1270°C
- reducing conditions	1290°C
Vanadium in ash	2.2 %
Nickel in ash	0.7 %

Moisture, ash, carbon, sulphur, calcium, sodium, vanadium and nickel contents and particle sizes were analysed for each test run. The results of these analyses are given in Appendix 1.

Average values for sulphur, vanadium and nickel in the coke (from the data in Appendix 1) were:

- sulphur	6.75 %
- vanadium	0.20 %
- nickel	0.066 %

6.2 Limestone

Loss on ignition, moisture, screen analysis, carbon, sulphur, calcium, magnesium, carbonate and chloride were determined for an overall sample divided into three samples. In addition, calcium, sodium, potassium, vanadium and nickel were analysed from a composite sample for each test run. The results are given in Appendix 2.

It was noticed during the tests that fly ash was easily agglomerated and quite difficult to convey pneumatically. The limestone contains 1-2% of alkaline compounds which were concentrated in the fly ash; fly ash contains 2-3% of alkaline compounds.

6.3 Process data

The process data were collected by an ALCONT automatic process control system. Every 5 minutes the computer calculated the mean values, scaled the channel voltages to the engineering units and gave a print out. A similar calculation and print out of mean values was performed every 30 minutes. A summary of process data for the test runs is presented in Table 3, which is given in the following pages. The notes below apply to the data in Table 3.

NOTES TO TABLE 3:

Note 1: Coke feed rate is expressed on an 'as-fired' basis, i.e. containing 3-3.5% moisture.

Note 2: Total air masses are as measured by calibrated orifice meters.

Note 3: Air coefficient is a factor defined as:

$$\text{Air coefficient} = \frac{\text{Total measured combustion air}}{\text{Calculated stoichiometric air}}$$

Note 4: 'Baghouse ash + particulates' is the sum of the ash collected and weighed in the baghouse ash hopper, plus the calculated weight of fly ash escaping the baghouse and exhausted to the stack (determined from isokinetic sampling of the baghouse exhaust gases).

Note 5: Bed variation is calculated from the measured change in bed inventory for each test run divided by the duration of each test run.

Note 6: Pressure under grid is the mean air pressure under the bed support grid during the steady-state test period.

Table 3. Process data

Run No.		0(Scoping)	1	2	3
Date		Dec 05	Dec 10	Dec 11	Dec 12
Time		0300-0900	0500-0800	0300-1400	1200-0900
Coke feed (1)	g/s	31.0	34.9	32.1	36.4
Dry solids (d.s.)	%	95.6	95.5	96.2	96.9
Ash in d.s.	%	8.30	9.44	7.73	7.89
S in d.s.	%	6.72	6.64	6.78	6.56
Limestone feed	g/s	0	0	0	0
Air, total (2)	g/s	340	375	354	425
Primary air	g/s	214	236	227	272
Secondary air	g/s	126	139	127	153
Air coefficient (3)		1.16	1.15	1.15	1.21
Flue gas	g/s	367	404	382	458
O_2 (in dry gas)	%	3.5	3.6	3.2	3.8
CO (in dry gas)	ppm	150	650	110	80
SO_2 (in dry gas)	ppm	n/a	4300	4800	5000
NOx (in dry gas)	ppm	220	100	150	520
Temperatures					
0.35 m	°C	891	810	889	930
1.50 m		917	840	917	955
2.50 m		907	830	908	952
3.50 m		900	830	892	940
7.00 m		882	810	864	926
8.50 m		879	815	872	935
Cyclone		894	830	889	941
After the boiler		253	258	300	307
Ash, total	g/s	3.0	5.3	3.4	3.1
Bottom ash	g/s	0	0	0	0
Boiler ash	g/s	2.5	1.3	1.0	0.6
Baghouse ash (4) + particulates	g/s	1.8	1.4	2.5	2.7
Bed variation (5)	g/s	-0.8	2.0	-1.0	-0.6
Accumulation + error	g/s	-0.5	0.6	0.9	0.4
Baghouse ash recycle	%	0	0	0	0
Pressure under grid	mbar	66	65	54	75
Initial bed mass	kg	302	288	291	329
Final bed mass	kg	285	310	250	311

Table 3. Process data (continued)

Run No.		4	5	6	7A
Date		Dec 17	Dec 18	Jan 7	Dec 20
Time		0300-1200	1430-2300	0530-1345	1000-0730
Coke feed (1)	g/s	34.6	36.9	36.3	33.0
Dry solids (d.s.)	%	97.1	97.4	96.9	97.5
Ash in d.s.	%	6.97	6.84	7.43	6.85
S in d.s.	%	6.80	6.64	6.64	6.72
Limestone feed	g/s	20.1	15.4	21.3	20.4
Air, total (2)	g/s	434	424	410	383
Primary air	g/s	269	267	242	242
Secondary air	g/s	175	157	168	142
Air coefficient (3)		1.29	1.18	1.17	1.19
Flue gas	g/s	467	457	444	414
O_2 (in dry gas)	%	4.7	3.2	3.4	3.5
CO (in dry gas)	ppm	50	70	90	120
SO_2 (in dry gas)	ppm	480	500	480	460
NOx (in dry gas)	ppm	310	130	200	160
Temperatures					
0.35 m	°C	925	867	845	822
1.50 m		955	895	861	847
2.50 m		951	894	840	840
3.50 m		933	878	856	820
7.00 m		923	871	858	827
8.50 m		937	880	864	839
Cyclone		948	891	873	847
After the boiler		274	339	317	321
Ash, total	g/s	21.2	18.4	22.5	21.0
Bottom ash	g/s	9.3	10.6	14.7	10.8
Boiler ash	g/s	1.1	0.5	0.7	1.4
Baghouse ash (4) + particulates	g/s	10.9	6.1	5.4	7.7
Bed variation (5)	g/s	0.8	-0.3	2.3	-0.6
Accumulation + error	g/s	-0.9	1.5	-0.6	1.7
Baghouse ash recycle	%	0	0	0	0
Pressure under grid	mbar	84	92	80	85
Initial bed mass	kg	344	422	326	445
Final bed mass	kg	370	414	394	425

Table 3. Process data (continued)

Run No.		8A	7B	8B	9
Date		Dec 19	Jan 9-10	Jan 10	Jan 10
Time		0650-1445	2030-0300	0630-1250	1430-2100
Coke feed (1)	g/s	35.1	32.6	34.3	36.1
Dry solids (d.s.)	%	97.1	96.3	96.8	96.9
Ash in d.s.	%	7.21	6.65	7.10	6.76
S in d.s.	%	6.66	6.82	6.80	6.80
Limestone feed	g/s	16.9	14.7	15.0	17.6
Air, total (2)	g/s	433	412	421	414
Primary air	g/s	273	255	261	257
Secondary air	g/s	160	157	160	158
Air coefficient (3)		1.27	1.31	1.26	1.18
Flue gas	g/s	465	443	452	448
O_2 (in dry gas)	%	4.5	5.0	4.3	3.2
CO (in dry gas)	ppm	40	30	45	40
SO_2 (in dry gas)	ppm	490	470	480	596
NOx (in dry gas)	ppm	120	175	185	245
Temperatures					
0.35 m	°C	869	823	881	922
1.50 m		895	853	911	952
2.50 m		896	855	912	953
3.50 m		886	845	896	940
7.00 m		879	831	879	927
8.50 m		881	843	896	933
Cyclone		890	853	910	942
After the boiler		320	314	301	290
Ash, total	g/s	19.0	17.0	17.6	19.6
Bottom ash	g/s	11.9	14.6	10.5	9.8
Boiler ash	g/s	0.8	1.8	3.2	3.5
Baghouse ash (4) + particulates	g/s	4.9	2.7	1.6	7.1
Bed variation (5)	g/s	2.9	-0.9	0.6	-0.3
Accumulation + error	g/s	-1.5	-1.2	1.7	-0.5
Baghouse ash recycle	%	0	80	80	80
Pressure under grid	mbar	97	91	86	92
Initial bed mass	kg	347	312	298	325
Final bed mass	kg	430	290	311	318

Table 3. Process data (continued)

Run No.		10	11	12	13
Date		Jan 15	Jan 15	Jan 14-15	Jan 15
Time		1530-1900	2100-2400	2015-0030	0100-0750
Coke feed (1)	g/s	38.9	40.9	24.5	23.2
Dry solids (d.s.)	%	96.7	96.3	96.6	96.5
Ash in d.s.	%	7.45	7.86	8.38	8.65
S in d.s.	%	6.89	6.77	6.88	6.83
Limestone feed	g/s	19.3	20.2	9.9	9.0
Air, total (2)	g/s	451	483	303	290
Primary air	g/s	280	300	179	174
Secondary air	g/s	171	184	125	116
Air coefficient (3)		1.20	1.24	1.30	1.32
Flue gas	g/s	487	521	325	310
O_2 (in dry gas)	%	3.5	4.0	4.8	5.0
CO (in dry gas)	ppm	20	5	20	10
SO_2 (in dry gas)	ppm	470	430	490	440
NOx (in dry gas)	ppm	240	450	70	65
Temperatures					
0.35 m	°C	927	946	875	881
1.50 m		961	977	906	911
2.50 m		961	976	906	911
3.50 m		951	961	887	887
7.00 m		939	943	861	850
8.50 m		943	955	862	851
Cyclone		952	970	866	858
After the boiler		329	299	209	206
Ash, total	g/s	21.9	23.0	12.3	11.5
Bottom ash	g/s	15.7	10.4	8.4	5.6
Boiler ash	g/s	1.1	2.5	1.8	1.0
Baghouse ash (4) + particulates	g/s	5.7	4.0	1.9	1.3
Bed variation (5)	g/s	-2.4	3.0	-0.8	1.7
Accumulation + error	g/s	1.8	3.1	1.0	1.9
Baghouse ash recycle	%	0	80	80	0
Pressure under grid	mbar	113	117	96	90
Initial bed mass	kg	425	414	320	300
Final bed mass	kg	395	446	308	342

Table 3. Process data (continued)

Run No.		14	15	16
Date		Jan 14	Jan 14	Jan 16-17
Time		0900-1305	1330-1800	0900-1000
Coke feed (1)	g/s	20.4	18.6	34.7
Dry solids (d.s.)	%	97.3	96.5	96.7
Ash in d.s.	%	7.22	8.25	7.71
S in d.s.	%	6.72	6.71	6.89
Limestone feed	g/s	6.4	5.0	14.7
Air, total (2)	g/s	239	210	411
Primary air	g/s	124	109	255
Secondary air	g/s	115	101	156
Air coefficient (3)		1.20	1.18	1.24
Flue gas	g/s	256	226	443
O_2 (in dry gas)	%	3.6	3.2	4.0
CO (in dry gas)	ppm	45	50	10
SO_2 (in dry gas)	ppm	580	520	480
NOx (in dry gas)	ppm	54	97	220
Temperatures				
0.35 m	°C	860	877	888
1.50 m		893	907	917
2.50 m		891	905	916
3.50 m		865	880	897
7.00 m		831	842	883
8.50 m		833	840	900
Cyclone		839	843	909
After the boiler		163	177	352
Ash, total	g/s	8.8	7.6	17.8
Bottom ash	g/s	0.9	6.0	9.0
Boiler ash	g/s	0.6	1.1	4.4
Baghouse ash (4) + particulates	g/s	1.7	2.7	3.9
Bed variation (5)	g/s	5.0	-2.4	0.6
Accumulation + error	g/s	0.6	0.2	-0.1
Baghouse ash recycle	%	0	80	67
Pressure under grid	mbar	107	95	97
Initial bed mass	kg	344	375	392
Final bed mass	kg	417	336	442

6.4 Material and Energy Balances

The analyses given in Tables 2 and 3, and in Appendices 1-6, were used in the mass and energy balance calculations. The coke higher heating value (HHV) from Table 2 (28.56 MJ/kg) was corrected for each test to account for the variation in HHV caused by variations in the ash content of the coke. The correction was made using a factor (k1) which is expressed as:

$$k1 = \frac{100 - W_{ash}}{100 - 9.5}$$

The denominator uses an ash content of 9.5%, as determined from the coke ultimate analysis (Table 2). W_{ash} is given in Table 3. For example, for test run No. 0, W_{ash} = 8.3% and k = 1.0133.

The HHV for the fuel used in each test, and thus the fuel energy input to the reactor, are calculated from:

Energy input = Coke feed rate x coke dry solids x 28.56 x k

Example: test run No. 0:

Energy input = (31.0 x 0.956 x 28.56 x 1.0133) kW
= 858 kW (see Table 6)

For the energy balance the ash was considered inert. The heat effects of limestone calcination and sulphation were taken into account. The energy balance was calculated on the basis of the temperature measurements and flow rates on the cooling side, as well as the energy input to the reactor.

A similar factor (k2) was used to adjust the C, N, H and O contents of the coke for each run. This factor is defined as:

$$k2 = \frac{100 - W_{ash} - W_s}{100 - 9.5 - 6.6}$$

The denominator uses an ash content of 9.5% and a sulphur content of 6.6%, as determined from the coke ultimate analysis. Values for W_{ash} and W_s are given in Table 3.

The stoichiometric combustion air and flue gas flows were calculated on the basis of the measured fuel input, the measured O_2, CO, SO_2, and NOx contents in the flue gas and the mean unburnt carbon in the ash. The latter was calculated as a weighted average of the analysed carbon content in the outcoming ash streams.

6.4.1 Overall material balance

The overall material balance is presented in Table 4.

Table 4 . Material balance

Run No.	Total input g/s	Coke g/s	Limestone g/s	Air g/s	Total output g/s	Flue gas g/s	Ash (by diff) g/s
0	371	31	0	340	371	367	4
1	410	35	0	375	410	404	6
2	386	32	0	354	386	382	4
3	461	36	0	425	461	458	3
4	489	35	20	434	489	467	22
5	476	37	15	424	476	457	19
6	467	36	21	410	467	444	23
7A	436	33	20	383	436	414	22
8A	485	35	17	433	485	465	20
7B	460	33	15	412	460	443	17
8B	470	34	15	421	470	452	18
9	468	36	18	414	468	448	20
10	509	39	19	451	509	487	22
11	544	41	20	483	544	521	23
12	338	25	10	303	338	325	13
13	322	23	9	290	322	310	12
14	265	20	6	239	265	256	9
15	234	19	5	210	234	226	8
16	461	35	15	411	461	443	18

It can be seen from Table 4 that the mass of the total ash was roughly four times greater when limestone was used for SO_2 reduction than when no limestone was used. The limestone feed rate is proportional to the coke feed rate and to the Ca:S molar ratio being used for each test.

6.4.2 Ash balance

The bottom, boiler and baghouse ash were measured by means of weighing hoppers hanging from strain gauges. The mass of bed material was calculated from a calibration curve relating bed mass to the air pressure under the bed support plate. The calibration curve is shown on Figure 7 (equation: $y = 4x + 50$). Ash balance data are presented in Table 5.

Figure 7. Bed material mass calibration curve

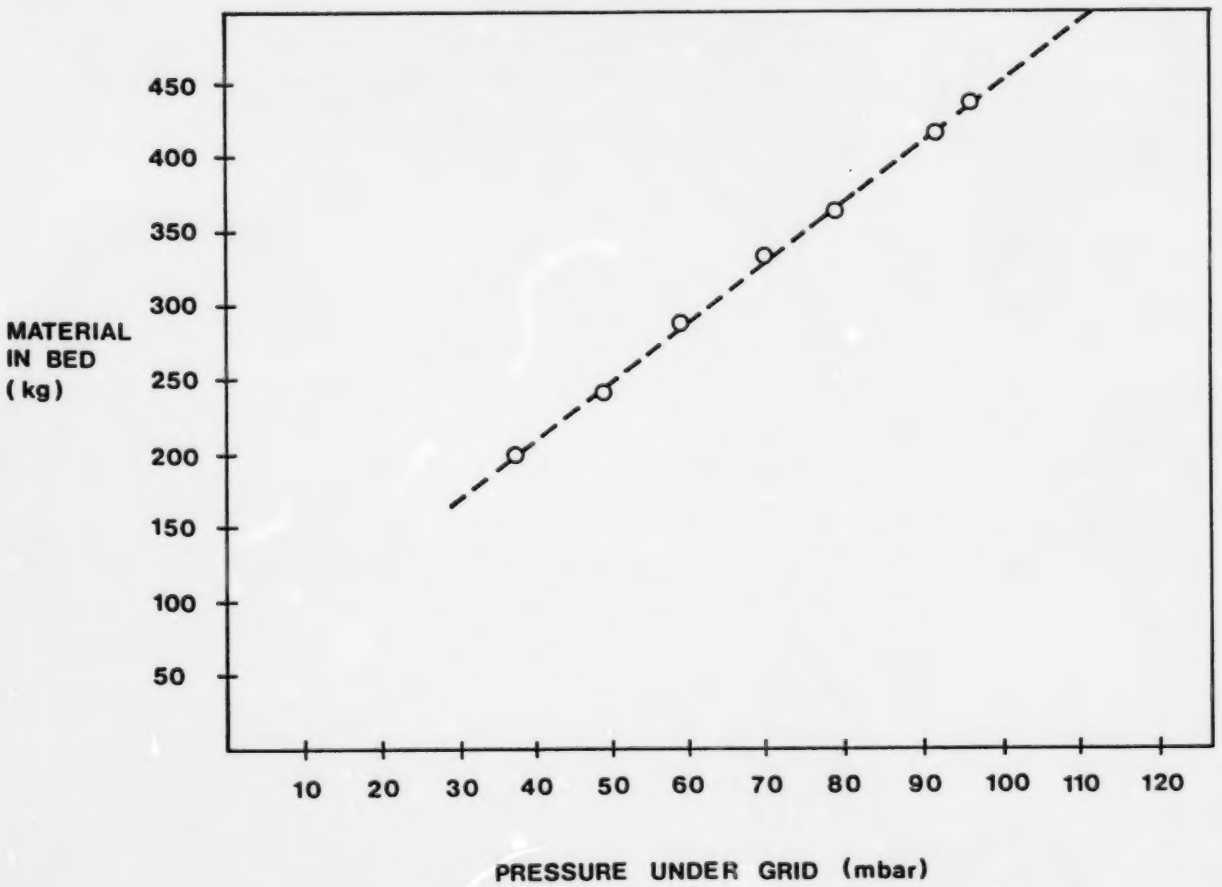


Figure 7 was used to derive the initial (M1) and final (M2) bed masses for each test run; these are given in Table 3. M1 and M2 are used in the elemental mass balances, as explained in Section 6.6.1, and in the overall ash balance (Table 5) to give the bed variation.

Table 5. Ash balance

Run No.	INPUT TOTAL g/s	Ash g/s	CaO g/s	$CaSO_4$ g/s	OUTPUT TOTAL g/s	Bottom ash g/s	Boiler ash g/s	Bag-house ash g/s	Parti-culate g/s	Bed Vari-ation g/s	Accum. + error g/s
0	3.0	3.0	-	-	3.0	0	2.5	1.8	.032	-0.8	-0.5
1	5.3	5.3	-	-	5.3	0	1.3	1.4	.016	2.0	0.6
2	3.4	3.4	-	-	3.4	0	1.0	-	2.5	-1.0	0.9
3	3.1	3.1	-	-	3.1	0	0.6	-	2.7	-0.6	0.4
4	21.2	7.9	4.6	8.7	21.2	9.3	1.1	10.9	.029	0.8	-0.9
5	18.4	6.6	2.7	9.1	18.4	10.6	0.5	6.1	.026	-0.3	1.5
6	22.5	7.9	5.5	9.1	22.5	14.7	0.7	5.4	n/a	2.3	-0.6
7A	21.0	8.1	4.5	8.4	21.0	10.8	1.4	7.7	n/a	-0.6	1.7
8A	19.0	7.1	3.3	8.6	19.0	11.9	0.8	4.9	.040	2.9	-1.5
7B	17.0	6.1	2.7	8.2	17.0	14.6	1.8	2.7	.027	-0.9	-1.2
8B	17.6	6.0	3.0	8.6	17.6	10.5	3.2	1.6	n/a	0.6	1.7
9	19.6	7.0	3.4	9.2	19.6	9.8	3.5	7.1	.041	-0.3	-0.5
10	21.9	8.1	3.7	10.1	21.9	15.7	1.1	5.7	.018	-2.4	1.8
11	23.0	8.7	3.9	10.4	23.0	10.4	2.5	4.0	.026	3.0	3.1
12	12.3	4.8	1.3	6.2	12.3	8.4	1.8	1.9	.016	-0.8	1.0
13	11.5	4.2	1.4	5.9	11.5	5.6	1.0	1.3	.008	1.7	1.9
14	8.8	3.0	0.8	5.0	8.8	0.9	0.6	1.7	.006	5.0	0.6
15	7.6	2.7	0.3	4.6	8.8	6.0	1.1	2.7	n/a	-2.4	0.2
16	17.8	6.4	2.5	8.9	17.8	9.0	4.4	3.9	.021	0.6	-0.1

The ash figure is the sum of the ash in the coke feed, the limestone residue (after ignition), and the carbon in the ash streams (as given in Table 7). The CaO is the residual unreacted CaO from the limestone. The $CaSO_4$ is calculated from the amount of CaO converted to $CaSO_4$, as determined from the sulphur capture for each test (Table 8).

The baghouse was by-passed in runs 2 and 3, because of a mechanical problem, and fly ash was removed in a venturi scrubber. Because the particulate isokinetic sampler is located between the baghouse and the venturi scrubber (see Figure 6) the particulate loadings for runs 2 and 3 are much higher than for other runs.

No bottom ash was drained from the reactor in runs 0 to 3 that were carried out without limestone addition. Baghouse ash recirculation in runs 7B, 8B, 9, 11, 12, 15 and 16 raised the amount of boiler ash. Roughly half of the total ash was drained out as bottom ash when limestone was added.

Bed variation is calculated from the difference in bed inventory during the steady-state period of each test. A positive value indicates an increase in bed inventory and a negative value indicates a decrease in bed inventory.

Fly ash was easily agglomerated and accumulated upon the heat transfer surfaces in the convection section of the boiler. The probable reason for the stickiness of the fly ash is the high alkali content in the limestone.

6.4.3 **Energy balance**

The calculation of the heat input in the energy balance is based on the higher heating value of dry fuel (ASME Standard) determined from the ultimate analysis of coke (28.56 MJ/kg), corrected for variations in coke ash content, as described previously on page 29 (Section 6.4). The reference temperature is 0°C.

The calculation of the heat output in the energy balance is based on the flow and temperature measurements of the cooling water, flue gas and ash streams, and on analyses of carbon in ash streams and carbon monoxide in flue gas.

The results are given in Table 6.

Table 6. Energy balance

Run No.	INPUT kW	Coke kW	Lime-stone kW	Air kW	OUTPUT kW	Flue gas kW	Ash kW	Un-burned carbon kW	Cooling effect kW	Loss + error kW
0	865	858	-	7	865	108	1	36	640	80
1	961	953	-	8	961	121	1	32	667	140
2	907	900	-	7	907	131	1	28	681	66
3	1035	1026	-	9	1035	159	1	15	723	137
4	1003	987	7	9	1003	145	3	14	702	139
5	1080	1057	14	9	1080	175	3	17	732	153
6	1041	1028	5	8	1041	160	5	39	735	100
7a	960	946	6	8	960	150	5	26	644	135
8a	1018	998	11	9	1018	168	4	16	697	133
7b	945	926	11	8	945	157	4	17	634	133
8b	994	974	12	8	994	154	5	9	679	147
9	1050	1030	12	8	1050	148	3	16	733	150
10	1121	1099	13	9	1121	181	4	13	809	114
11	1167	1146	11	10	1167	177	4	18	789	179
12	702	685	11	6	702	79	1	7	408	207
13	662	646	10	6	662	74	1	6	385	196
14	596	581	10	5	596	50	1	10	388	147
15	534	520	10	4	534	48	1	5	391	89
16	998	976	14	8	998	160	3	13	709	113

The flue gas enthalpy is quite high because of the high temperature after the boiler. Heat losses due to ash and unburned carbon are quite low.

6.4.4 Combustion efficiency

The calculation of combustion efficiency included:

- the total fuel mass entering the reactor and the energy contained in the fuel, based on its corrected higher heating value.
- the average carbon content of the baghouse, boiler and bottom ash streams and their relative mass flow rates.
- the carbon monoxide (CO) content of the flue gases.

The energy losses in unburned carbon and CO were calculated using the following values:

$C + O_2 = CO_2$ 33.73 MJ/kg C
$CO + O = CO_2$ 0.283 MJ/mole CO

Losses and calculated combustion efficiencies are presented in Table 7.

Table 7. Combustion efficiency

Run No.	Carbon in bottom ash		Carbon in boiler ash		Carbon in baghouse ash		CO in dry flue gas		Energy input	Combustion efficiency	Baghouse ash recycle
	%	kW	%	kW	%	kW	ppm	kW	kW	%	%
0	-	-	22.5	19.0	21.3	12.9	150	0.5	858	96.2	0
1	-	-	26.1	11.4	23.7	11.2	650	2.4	953	97.4	0
2	-	-	19.3	6.5	*16.1	13.6	110	0.4	900	97.7	0
3	-	-	10.9	2.2	*11.1	10.1	80	0.3	1026	98.8	0
4	0.6	1.9	4.5	1.7	4.0	14.5	50	0.2	987	98.1	0
5	0.8	2.9	6.4	1.1	5.5	11.4	70	0.3	1057	98.5	0
6	0.9	4.5	8.8	2.1	8.3	15.2	90	0.4	1028	97.8	0
7a	1.1	4.0	5.8	2.7	4.8	12.4	120	0.5	946	97.9	0
8a	0.8	3.1	6.2	1.7	6.0	9.9	40	0.2	998	98.5	0
7b	2.4	11.7	6.4	3.9	3.5	3.2	30	0.1	926	98.0	80
8b	1.0	3.5	4.4	4.7	2.0	1.1	45	0.2	974	99.0	80
9	0.5	1.7	3.0	3.5	1.8	4.4	40	0.2	1030	99.0	80
10	1.9	6.9	3.8	1.4	3.8	7.3	20	0.1	1099	98.6	0
11	1.9	6.7	3.0	2.5	3.1	4.2	5	0.0	1146	98.8	80
12	1.0	2.9	3.3	2.8	1.6	1.0	20	0.1	685	99.0	80
13	0.9	1.7	4.9	1.6	1.8	0.8	10	0.0	646	99.4	0
14	0.8	0.2	5.8	1.2	3.0	1.7	45	0.1	581	99.4	0
15	0.9	1.8	5.2	1.9	1.2	1.1	50	0.1	520	99.1	80
16	1.3	4.1	3.9	5.7	2.9	3.7	10	0.0	976	98.6	67

* measured from particulate samples, baghouse bypassed

The combustion efficiency ranged from 96.2 to 99.4% over all the test runs. However, run No. 0 (96.2% efficiency) was a scoping test, when combustor operating conditions were being established for the tests and efficiency was not a major concern. Eliminating this run, the combustion efficiency ranged from 97.4 to 99.4%, with an average of 98.6%.

It is noticeable that as the test program progressed, and more experience was gained with burning the fuel, the combustion efficiency increased. For test runs 8B to 16, the combustion efficiency ranged from 98.6 - 99.4%, with an average of 99%.

Regression analysis was used to calculate the effect on combustion efficiency of the following variables:

1. Combustion efficiency was the response variable (%)
2. Average reactor temperature (^{o}C)
3. Superficial gas velocity in freeboard (m/s)
4. Excess O_2 in dry flue gas (%)
5. Primary air (% of total combustion air)
6. Fly ash recirculation from baghouse (% of ash coming into baghouse)

Variables 2-6 were factor variables.

The proportion of primary air has the greatest effect on combustion efficiency. Baghouse ash recirculation improves combustion efficiency to some extent. Reactor temperature and excess O_2 also seem to have a minor effect.

The combination of the variables primary air, temperature and excess O_2 gives the smallest standard deviation and the highest F value. The variation for the different variables is, however, small and the relationship is not very conclusive.

6.5 Emissions

6.5.1 SO_2 emissions

SO_2 reduction without limestone addition varied from 0 to 19.7%. The inherent sulphur capture by the coke ash and bed material decreases with increasing temperature. At 940°C all the sulphur in the coke is released as SO_2. With limestone addition the target was to reduce the SO_2 emission by 90%. The SO_2 retention varied between 88.6 and 91.4%.

The SO_2 reduction data are given in Table 8.

Table 8. SO_2 reduction data

Run No.	SO_2 ppm	S capture %	Ca:S (1)	Mean temp. °C(2)	Ash recirc. %	SFV (3) m/s	Comment
0	n/a	-	-	-	-	4.6	Scoping test
1	4300	19.7	0	823	0	4.7	No limestone
2	4800	10.4	0	890	0	4.7	No limestone
3	5000	0.0	0	940	0	5.9	No limestone
4	480	89.9	2.04	937	0	6.1	
5	500	90.2	1.54	881	0	5.6	
6	480	90.7	2.26	854	0	5.3	
7A	460	91.0	2.11	833	0	4.9	
8A	490	89.6	1.71	884	0	5.3	
7B	470	90.0	1.64	842	80	5.3	
8B	480	90.1	1.65	896	80	5.6	
9	490	90.6	1.72	938	80	5.8	
10	470	90.9	1.73	947	0	6.3	Max. load
11	430	91.4	1.76	960	80	6.9	Max. load
12	490	89.9	1.34	883	80	4.0	
13	440	90.8	1.41	882	0	3.8	
14	580	88.6	1.21	862	0	3.1	Min load
15	520	90.0	1.03	875	80	2.8	Min load
16	480	90.5	1.53	900	67	5.5	Optimum

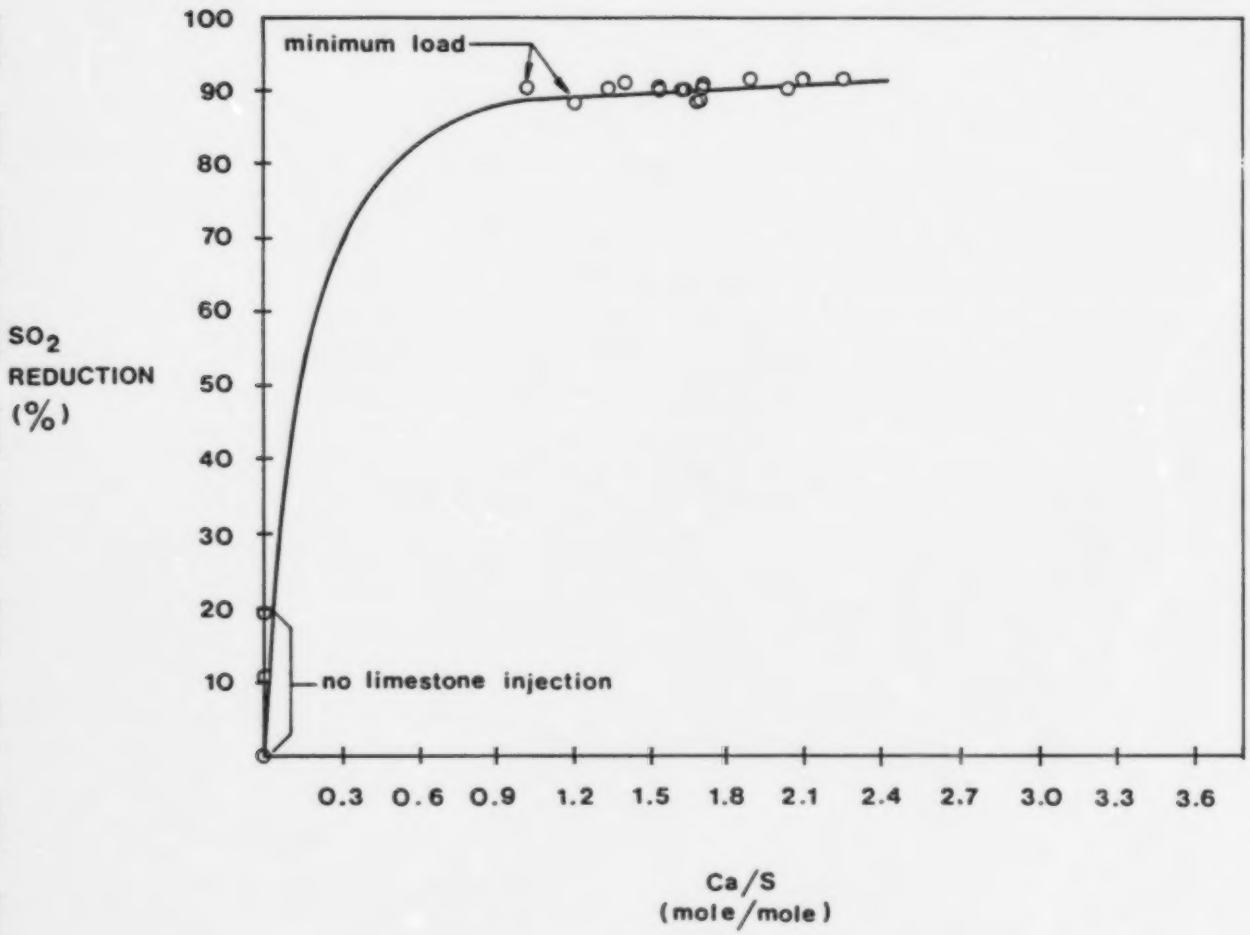
Notes to Table 8:

(1) Ca:S is moles calcium in limestone: moles of sulphur in coke
(2) Mean reactor temperature
(3) Superficial Fluidising Velocity in the reactor freeboard

One of the key test parameters was the reduction of SO_2 emissions by 90%. As can be seen from Figure 8, this was achieved in all tests with limestone injection. Sulphur retention varied between 88.6 and 91.4%, with a mean retention of 90%. With no limestone injection, sulphur retention varied from 0 to 19.7%.

Since, however, the amount of SO_2 retention was a constraint on the test program, the effect of differing Ca:S molar ratios on sulphur capture is masked. The level of sulphur retention is proportional to the amount of limestone used, but only over a narrow range. This is illustrated in the next three figures.

Figure 8. Sulphur capture Vs Ca:S molar ratio for all tests



As discussed previously, the constraint that sulphur capture should be 90% masks the effect of different factors on sulphur capture. In particular, the effect of Ca:S molar ratio is not sharply defined. Figures 9 and 10 do indicate that increasing the Ca:S molar ratio reduces SO_2 emissions, but not to the degree anticipated.

Figure 9 indicates the degree by which sulphur capture and limestone utilisation efficiency are affected by continuity in the test program, i.e. as the program progresses, and the pilot-plant staff become more familiar with the fuel, they are able to achieve 90% sulphur capture with steadily decreasing Ca:S molar ratios.

Figure 9. Sulphur capture & Ca:S molar ratio variation

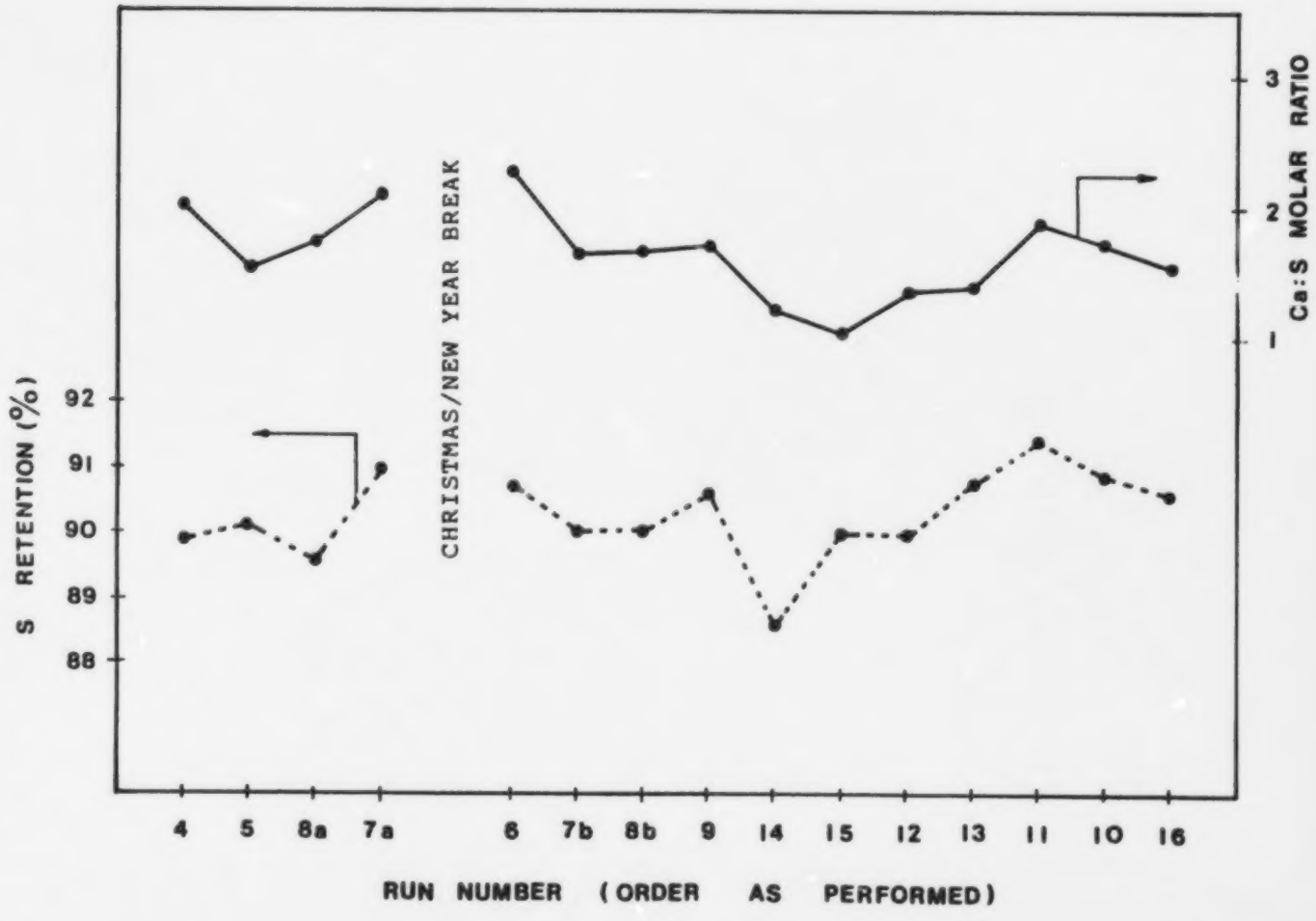
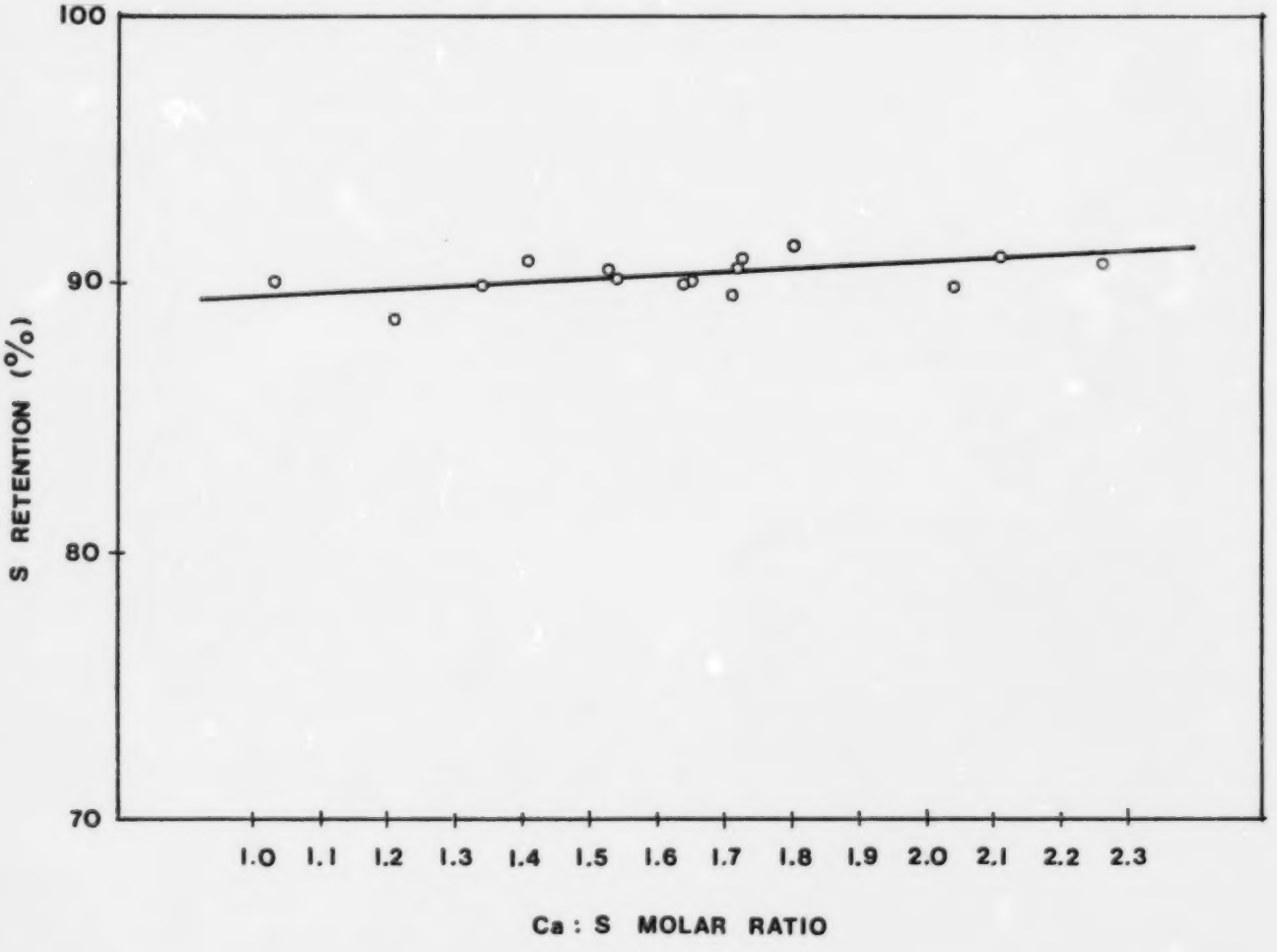


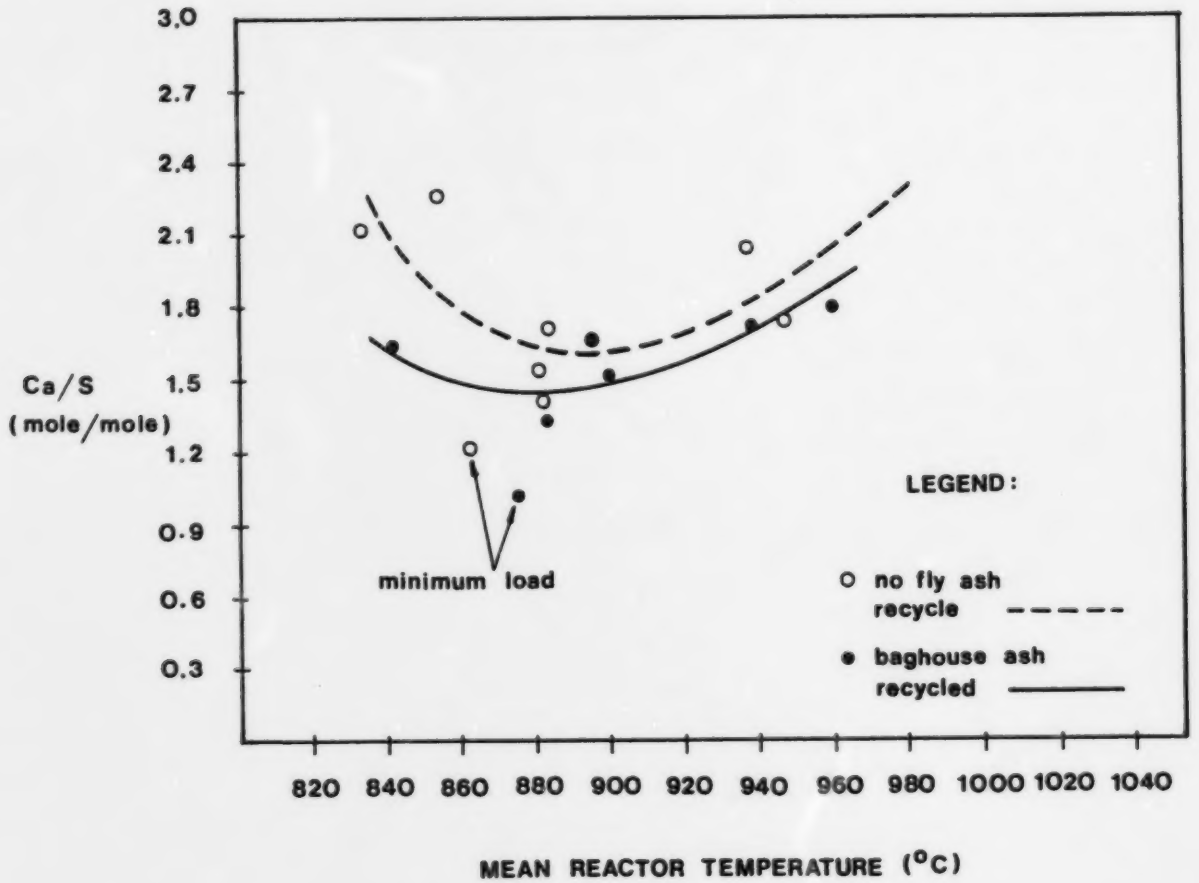
Figure 10 indicates that as Ca:S molar ratio increases there is a gradual improvement in sulphur capture. As noted previously, the improvement is not as marked as would be expected from the increase in Ca:S molar ratio from 1.03 to over 2.2. The 90% sulphur retention target prevented a detailed exploration of the effect of varying Ca:S molar ratio on sulphur capture.

Figure 10. Sulphur capture Vs Ca:S molar ratio (expanded scale)



The effect of reactor temperature on Ca:S molar ratio is shown in Figure 11. As can be seen, there is an optimum temperature of about 900°C for SO_2 reduction. Baghouse ash recirculation seems to reduce the Ca:S molar ratio by about 10% for temperatures over about 900°C and by somewhat more (up to 20%) for temperatures between 840 and 880°C. With minimum load, the SO_2 reduction is more efficient due to a longer residence time in the reactor. The optimum temperature range varies between 860 and 900°C for obtaining the lowest Ca:S molar ratio.

Figure 11. Ca:S molar ratio Vs reactor temperature



In order to get information about the parameters which have the greatest influence on SO_2 retention, the data were analysed by multivariable regression analysis. The following variables were chosen:

1. Sulphur retention (%) was the response variable.
2. Ca:S molar ratio (mole/mole). Calcium inherent in coke ash was not included.
3. Average reactor temperature (^oC)
4. Superficial gas velocity in freeboard (m/s)
5. Excess O_2 in dry flue gas (%)
6. Primary air proportion of total combustion air (%)
7. Fly ash recirculation rate (%)

The analysis showed that the Ca:S molar ratio has the greatest effect on sulphur retention. Baghouse ash recirculation and using more primary air improve sulphur retention. The effects of reactor temperature and superficial gas velocity seem to be minor. However, there seems to be an optimum temperature of about 900°C (Figure 11) that did not show up in the regression analysis.

6.5.2 **NOx emissions**

The measured NOx emissions varied from 54 to 520 ppm (in dry gas) corresponding to 0.035 - 0.341 mg NO_2/kJ. The effect of the following variables on NOx emissions were studied by regression analysis:

1. NOx emission (mg NO_2/kJ) was the response variable
2. Primary air, as a percentage of total combustion air (%)
3. Mean reactor temperature (^oC)
4. CO content in dry flue gas (ppm)
5. Excess O_2 in dry flue gas (%)
6. Fly ash recirculation rate (%)
7. SO_2 emission and limestone feed rate

The NOx emissions for each test are presented in Table 9, and are shown on Figure 12. It is apparent, for the range of sulphur retention, excess air and baghouse ash recirculation rates examined during this series of tests, that these parameters have very little influence upon NOx emission levels. The purpose of the tests was to provide optimum operating conditions for the fuel, and to enable boiler design conditions to be established. The test levels were therefore pre-set to encompass normal boiler operating conditions rather than an in-depth study of NOx emissions. For example, sulphur retention requirements were set at 90%, so we are unable to correlate SO_2 and NOx.

The test series demonstrated that by careful design and selection of the operating parameters, NOx emission levels as low as 50-60 ppm (0.032-0.039 mg/kJ) can be achieved. As would be expected, the NOx emission levels tend to increase as the combustor temperature increases. This effect is shown on Figure 13.

Figure 12. NOx emissions for each test

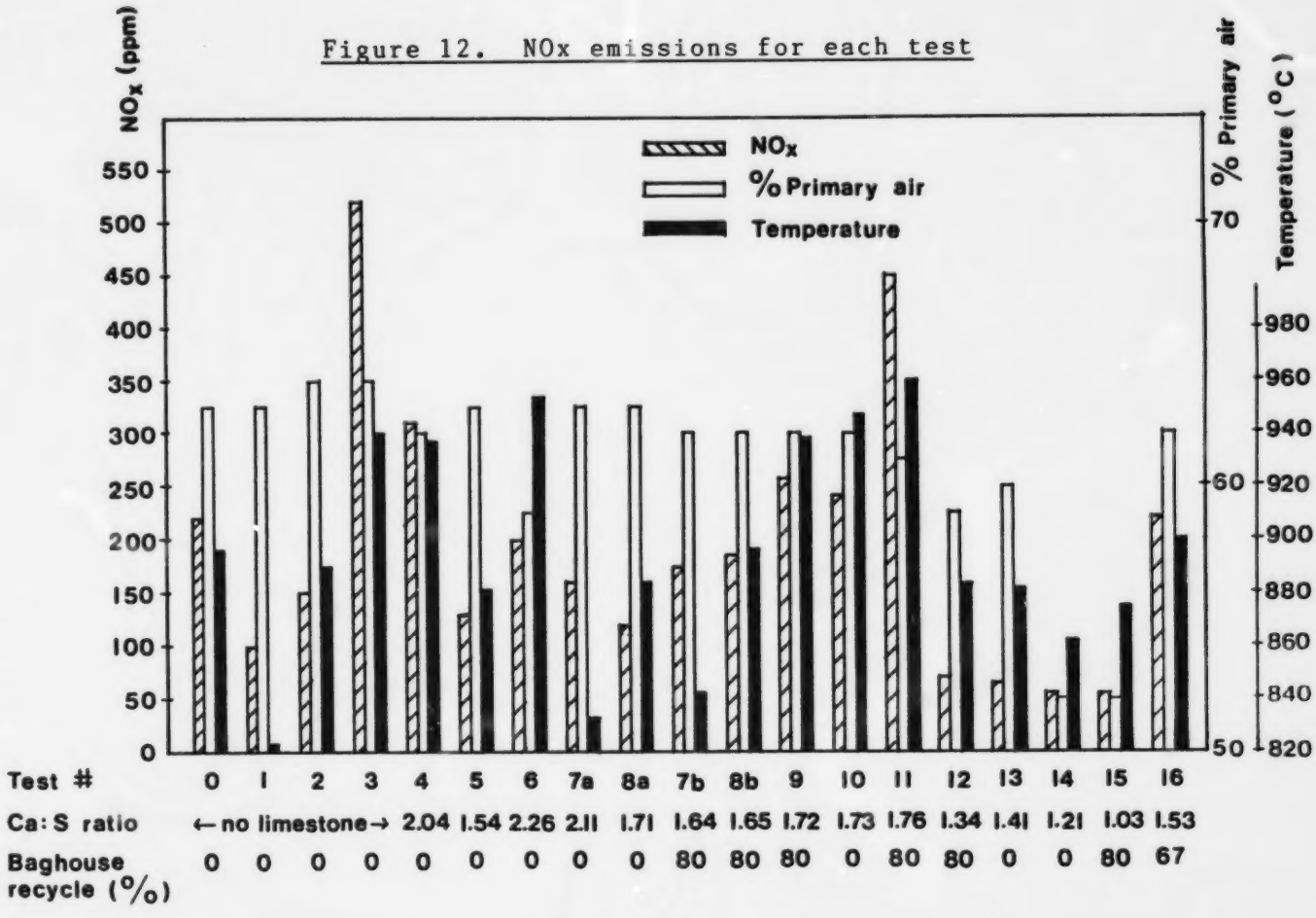


Figure 13. NOx emissions Vs combustor temperature

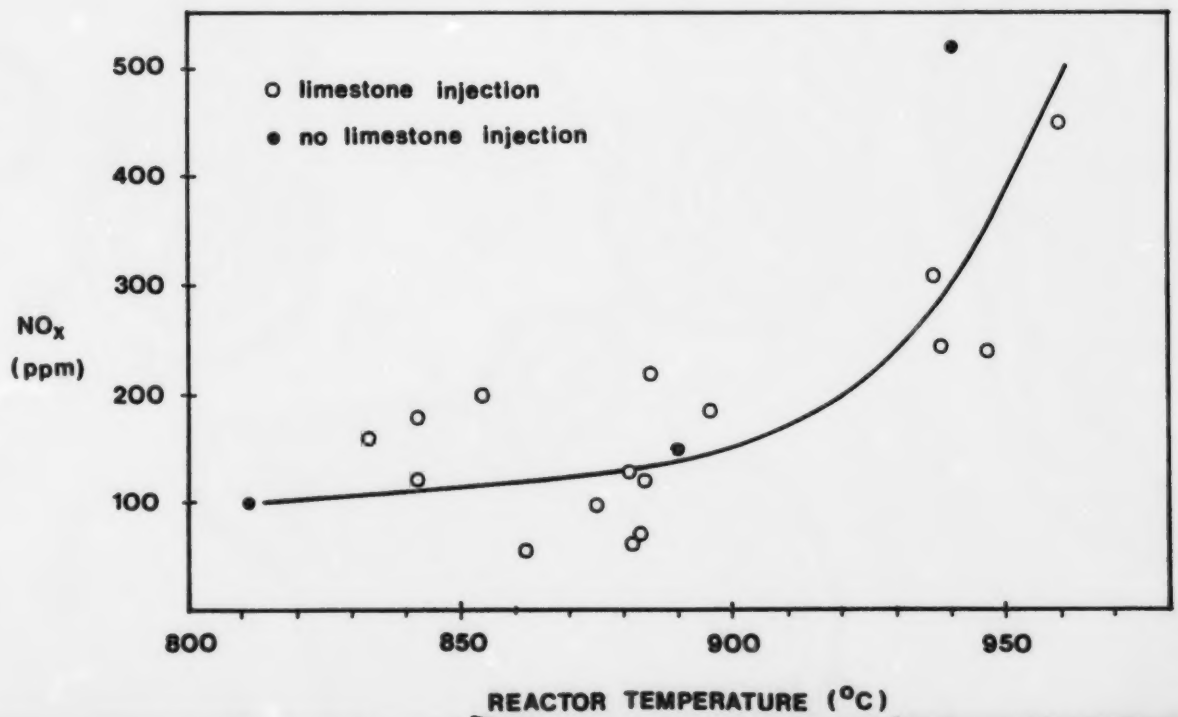


Table 9. NOx emission data

Run No.	NOx ppm	Temp °C	O_2 %	Ash recirc. %	CO ppm	SO_2 ppm	Ca:S molar ratio	P:A (1) %	Note
0	220	896	3.5	0	150	n/a	-	63	Scoping
1	100	823	3.6	0	650	4300	-	63	No
2	150	890	3.2	0	110	4800	-	64	lime-
3	520	940	3.8	0	80	5000	-	64	stone
4	310	937	4.7	0	50	480	2.04	62	
5	130	881	3.2	0	70	500	1.54	63	
6	200	954	3.4	0	90	480	2.26	59	
7A	160	833	3.5	0	120	460	2.11	63	
8A	120	884	4.5	0	40	490	1.71	63	
7B	175	842	5.0	80	30	470	1.64	62	
8B	185	896	4.3	80	45	480	1.65	62	
9	254	938	3.2	80	40	490	1.72	62	
10	240	947	3.5	0	20	470	1.73	62	Max.
11	450	960	4.0	80	5	430	1.76	61	load
12	70	883	4.8	80	20	490	1.34	59	
13	65	882	5.0	0	10	440	1.41	60	
14	54	862	3.6	0	45	580	1.21	52	Min.
15	52	875	3.2	80	50	520	1.03	52	load
16	220	900	4.0	67	10	480	1.53	62	Optimum

Note(1): P:A refers to the proportion of primary air as a percentage of the total combustion air. From the above figures, the amount of secondary (or overfire) air is (100 - P:A).

For the levels of primary air investigated, it is apparent that increasing the primary/secondary air proportion causes NOx levels to increase. This is to be expected, since an increase in primary air will tend to raise the oxygen content within the lower portion of the reactor, causing the effect of staged combustion to be reduced, and increasing NOx formation. It may be seen from the NOx results that optimum NOx control occurs at an operating temperature of 840-880°C. The fuel is a low volatile fuel, however, and because of this a higher operating temperature (900°C) may be preferred. This will ensure good combustion stability with Syncrude coke; a fuel that is relatively difficult to ignite. No combustion problems were encountered in the tests, even at the lowest temperatures used.

6.5.3 **Particulate emissions**

Particulates were measured by isokinetic sampling of flue gas after the baghouse through a Kaowool filter. The baghouse was by-passed during runs 2 and 3, because of mechanical problems. As discussed previously, since the particulate sample is taken from the baghouse exhaust duct, the particulate measurements for runs 2 and 3 show high readings.

Calcium, sodium, vanadium and nickel were determined from the particulate samples. The analyses are given in Appendix 8. The measured particulate emissions and the amounts of calcium, sodium, vanadium and nickel are given in Table 10.

Table 10. Particulate emissions

Run No.	Particulates		Calcium		Sodium		Vanadium		Nickel	
	mg/Nm^3	mg/s	mg/s	% of input	mg/s	% of input	mg/s	% of input	mg/s	% of input
0	79	32	0.8	1	0.3	2	0.7	1	0.3	1.7
1	32	16	0.3	1	0.1	0.3	0.3	0.5	0.2	1.0
2	8900	2500	55	125	21	135	43	68	15	81
3	8200	2700	62	150	18	74	46	62	13	68
4	55	29	2.8	0.05	0.2	0.2	0.1	0.1	0.02	0.1
5	41	26	3.1	0.07	0.1	0.2	0.1	0.1	0.04	0.2
6	n/a	n/a	n/a		n/a		n/a		n/a	
7A	n/a	n/a	n/a		n/a		n/a		n/a	
8A	58	40	4.2	0.09	0.2	0.2	0.3	0.5	0.1	0.4
7B	44	27	2.2	0.05	0.08	0.1	0.2	0.3	0.08	0.4
8B	n/a	n/a	n/a		n/a		n/a		n/a	
9	66	41	3.5	0.07	0.2	0.2	0.3	0.4	0.1	0.4
10	19	18	1.8	0.03	0.1	0.1	0.1	0.1	0.03	0.1
11	35	26	2.5	0.04	0.08	0.08	0.05	0.06	0.02	0.07
12	40	16	1.2	0.04	0.04	0.09	0.05	0.1	0.02	0.1
13	22	8	0.6	0.02	0.05	0.1	0.05	0.1	0.02	0.1
14	25	6	0.5	0.02	0.02	0.05	0.02	0.05	0.005	0.03
15	n/a	n/a	n/a		n/a		n/a		n/a	
16	29	21	2.8	0.06	0.2	0.2	0.2	0.3	0.04	0.2

Less than 1% of the total calcium, sodium, vanadium and nickel in the fuel were emitted from the baghouse to the stack during tests using limestone for sulphur capture.

6.6 Sulphur, calcium, sodium, vanadium and nickel balances

6.6.1 Explanatory note

The mass balances for the different elements include the total mass input to the reactor in the coke and limestone, and the total mass output from the pilot-plant in the bottom, boiler and baghouse ash streams. Particulates, as measured by isokinetic sampling after the baghouse, are included in the baghouse ash total.

One other factor that must be included in an accurate mass balance is the variation in bed inventory and in elemental concentrations in the bed material during each test run. As shown in Table 5, the bed variation may be positive or negative, depending on whether the bed inventory increased or decreased during a test run.

The change in bed inventory of a given element for a test run is given by:

$$\text{Change in inventory} = M_1 \times C_1 - M_2 \times C_2$$

where M_1 = initial bed mass (kg) (Table 3)
M_2 = final bed mass (kg) (Table 3)
C_1 = initial bed concentration (Appendix 6)
C_2 = final bed concentration (Appendix 6)

If we let $M_1 - M_2 = m$ (change in bed mass over a run)
and let $C_1 - C_2 = c$ (change in concentration over a run)

Then:

$$\text{Change in inventory} = M_1 \times c + m(C_1 - c)$$

Since m and c may be either positive or negative, according to increases or decreases in the bed mass or concentration, the variations in elemental masses in the bed may also be either positive or negative. However, it should be noted that the bed elemental variations (+ or -) may not necessarily coincide with the variations in bed mass reported in Table 5.

For example: test run No.7B

Bed mass variation = -0.9 g/s (see Table 5)
Bed S variation = -0.07 g/s (see Table 11)
Bed Ca variation = -0.43 g/s (see Table 12)
Bed Na variation = + 0.0042 g/s (see Table 13)
Bed V variation = + 0.0060 g/s (see Table 14)
Bed Ni variation = + 0.0016 g/s (see Table 15)

The explanation for the positive values for Na, V and Ni is that although the total bed mass decreased, the concentration of these elements in the bed material increased (see Appendix 6).

6.6.2 Sulphur balance

The sulphur balances for all test runs are presented in Table 11. According to Table 11, 'Accumulation + error' account for about 30% inaccuracy. Difficulties in obtaining representative samples for chemical analysis and inaccuracies in weighing the ash streams (which generally had small mass flow rates) may explain some part of the error. Ash hold-up in the system, because of its relatively sticky nature, may also be a partial explanation.

With limestone addition, roughly half of the sulphur is captured by bottom ash, except for test run No. 14. During this test, the bed inventory was increased and bottom ash was not drained from the reactor. Boiler ash represents 2-16% of the captured sulphur and the rest (4-28%) is found in the baghouse ash.

Table 11. Sulphur balance

Run No.	INPUT Total	S in coke		S in limestone		OUTPUT Total	S in bottom ash		S in boiler ash		S in bag-house ash		S variation in bed		S in flue gas		Accumulation + error	
	g/s	g/s	% of input	g/s	% of input	g/s	g/s	% of input	g/s	% of input	g/s	% of input	g/s	% of input	g/s	% of input	g/s	% of input
0	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
1	2.22	2.22	100	-	-	2.22	-	-	0.04	2	0.04	2	0.02	1	1.78	80	0.34	15
2	2.09	2.09	100	-	-	2.09	-	-	0.02	1	0.07	3	0.00	0	1.87	90	0.13	6
3	2.31	2.31	100	-	-	2.31	-	-	0.01	0	0.06	3	-0.01	0	2.31	100	-0.06	-3
4	2.31	2.28	99	0.03	1	2.31	0.49	21	0.09	4	0.64	28	0.38	16	0.23	10	0.48	21
5	2.42	2.39	99	0.03	1	2.42	1.18	49	0.05	2	0.39	16	0.20	8	0.24	10	0.36	15
6	2.38	2.34	98	0.04	2	2.38	1.23	52	0.06	3	0.32	13	0.57	24	0.22	9	-0.02	-1
7a	2.20	2.16	98	0.04	2	2.20	1.24	56	0.12	5	0.51	23	-0.26	-12	0.20	9	0.39	19
8a	2.30	2.27	99	0.03	1	2.30	1.46	63	0.08	3	0.34	15	0.43	19	0.24	10	-0.25	-10
7b	2.17	2.14	99	0.03	1	2.17	1.33	61	0.15	7	0.18	8	-0.07	-3	0.22	10	0.36	17
6b	2.29	2.26	99	0.03	1	2.28	1.22	54	0.24	11	0.08	4	0.28	12	0.23	10	0.23	9
9	2.41	2.38	99	0.03	1	2.41	1.15	48	0.29	12	0.48	20	0.20	8	0.23	9	0.06	3
10	2.62	2.59	99	0.03	1	2.62	1.82	69	0.09	3	0.35	14	-0.24	-9	0.24	9	0.36	14
11	2.70	2.67	99	0.03	1	2.70	1.10	41	0.18	7	0.25	9	0.14	5	0.23	9	0.80	29
12	1.65	1.63	99	0.02	1	1.65	0.74	45	0.15	9	0.12	7	-0.07	-4	0.17	10	0.54	33
13	1.55	1.53	99	0.02	1	1.55	0.56	36	0.08	5	0.07	5	0.17	11	0.14	9	0.53	34
14	1.34	1.33	99	0.01	1	1.34	0.05	4	0.04	3	0.10	8	1.35	101	0.15	11	-0.35	-27
15	1.21	1.20	99	0.01	1	1.21	0.44	37	0.08	7	0.18	15	0.02	1	0.12	10	0.37	30
16	2.34	2.31	99	0.03	1	2.34	0.91	39	0.38	16	0.23	10	0.03	1	0.22	9	0.57	25

6.6.3 **Calcium balance**

As shown in Table 12, the distribution of calcium follows the distribution of sulphur in the solid ash streams.

In test runs 0 to 3 no limestone was added to the reactor, and the amounts of calcium in the system are very small, making for a large 'Accumulation + error'. In test run No. 14, the large change in bed inventory, when the bed mass was increased deliberately, makes the balance inexact and the 'Accumulation + error' figure is large.

Run 0 was a Scoping Test, during which the reactor was being set up and calibrated for the test runs. This, coupled with the extremely small quantities of calcium involved in the mass balance, may explain the very large percentage differences between the calcium input and the calcium in the ash streams.

Table 12 Calcium balance

Run No.	INPUT Total	Ca in coke		Ca in limestone		OUTPUT Total	Ca in bottom ash		Ca in boiler ash		Ca in baghouse ash		Ca variation in bed		Accumulation + error	
	g/s	g/s	% of input	g/s	% of input	g/s	g/s	% of input	g/s	% of input	g/s	% of input	g/s	% of input	g/s	% of input
0	0.01	0.01	100	-	-	0.01	-	-	0.03	300	0.04	400	0.00	0	-0.06	+600
1	0.03	0.03	100	-	-	0.03	-	-	0.01	33	0.02	67	0.01	33	-0.01	-33
2	0.04	0.04	100	-	-	0.04	-	-	0.01	25	0.05	125	-0.04	-100	0.02	50
3	0.04	0.04	100	-	-	0.04	-	-	0.01	25	0.06	150	-0.02	-50	-0.01	-25
4	5.90	0.078	1	5.83	99	5.90	2.16	37	0.19	3	1.94	33	0.77	13	0.84	14
5	4.67	0.08	2	4.59	98	4.67	2.40	51	0.10	2	1.22	26	0.12	3	0.83	18
6	6.68	0.08	1	6.60	99	6.68	3.90	58	0.15	2	1.02	15	0.96	14	0.65	11
7a	5.74	0.03	1	5.71	99	5.74	2.72	47	0.32	6	1.61	28	-0.19	3	1.28	16
8a	4.91	0.05	1	4.86	99	4.91	2.45	50	0.16	3	0.87	18	0.53	11	0.90	18
7b	4.42	0.04	1	4.38	99	4.42	3.46	78	0.52	12	0.49	11	-0.43	-10	0.38	9
8b	4.67	0.02	0	4.65	100	4.67	2.64	57	0.63	13	0.27	6	-0.09	-2	1.22	26
9	5.13	0.03	1	5.10	99	5.13	2.65	52	0.76	15	1.49	29	-0.44	-9	0.67	13
10	5.65	0.06	1	5.59	99	5.65	3.64	64	0.24	4	1.06	19	-0.53	-9	1.24	22
11	5.91	0.05	1	5.86	99	5.91	2.29	39	0.48	8	0.80	14	0.75	13	1.55	26
12	2.77	0.05	2	2.72	98	2.77	1.47	53	0.37	13	0.35	13	-0.20	-7	0.78	28
13	2.74	0.04	1	2.70	99	2.74	1.06	39	0.29	7	0.25	9	1.32	48	-0.08	-3
14	2.16	0.04	2	2.02	98	2.16	0.10	5	0.12	6	0.39	18	2.78	129	-1.23	-50
15	1.89	0.03	2	1.55	98	1.89	1.23	65	0.20	11	0.53	28	-0.34	-10	0.27	14
16	4.47	0.05	1	4.42	99	4.47	1.98	44	0.88	20	0.88	20	0.34	8	0.39	8

6.6.4 **Sodium balance**

Roughly 40% of the total sodium input is captured in the bottom ash, 10% in the boiler ash and 40% in the baghouse ash. The large change in bed inventory during test run No. 14 creates a large 'Accumulation + error' and affects the mass balance.

As shown in Table 13, the majority of the sodium input to the reactor (80-90%) is from the sodium contained in the limestone. As can be seen from the limestone analyses (Appendix 2), the limestone contained 0.35-0.44% sodium.

Table 13. Sodium balance

Run No.	INPUT Total	Na in coke		Na in limestone		OUTPUT Total	Na in bottom ash		Na in boiler ash		Na in baghouse ash		Na in variation in bed		Accumulation + error	
	g/s	g/s	% of input	g/s	% of input	g/s	g/s	% of input	g/s	% of input	g/s	% of input	g/s	% of input	g/s	% of input
0	0.0148	0.0148	100	-	-	0.0148	-	-	0.0178	120	0.0130	88	+0.0123	-83	+0.0037	-25
1	0.0334	0.0334	100	-	-	0.0334	-	-	0.0094	28	0.0127	38	0.0106	32	0.0007	2
2	0.0154	0.0154	100	-	-	0.0154	-	-	0.0092	60	0.0208	135	+0.0109	-71	+0.0037	-24
3	0.0247	0.0247	100	-	-	0.0247	-	-	0.0048	19	0.0184	74	+0.0067	-27	0.0082	34
4	0.0851	0.0067	8	0.0784	92	0.0851	0.0781	92	0.0077	9	0.0262	31	+0.0354	-42	0.0085	10
5	0.0646	0.0107	17	0.0539	83	0.0646	0.0233	36	0.0023	4	0.0293	45	0.0012	2	0.0085	13
6	0.0936	0.0106	11	0.0830	89	0.0936	0.0368	39	0.0034	4	0.0394	42	0.0010	1	0.0130	14
7a	0.0676	0.0064	9	0.0612	91	0.0676	0.0216	32	0.0067	10	0.0377	56	+0.0017	-3	0.0033	5
8a	0.0812	0.0178	21	0.0642	79	0.0812	0.0357	44	0.0021	3	0.0357	44	+0.0043	-5	0.0120	14
7b	0.0592	0.0063	11	0.0529	89	0.0592	0.0292	49	0.0079	13	0.0221	37	0.0042	7	+0.0042	-6
8b	0.0826	0.0166	20	0.0660	80	0.0826	0.0294	36	0.0131	16	0.0096	12	+0.0036	-4	0.0341	40
9	0.0826	0.0070	8	0.0756	92	0.0826	0.0343	42	0.0147	18	0.0490	59	+0.0015	2	+0.0139	-21
10	0.0788	0.0113	14	0.0675	86	0.0788	0.0314	40	0.0054	7	0.0285	36	+0.0046	-6	0.0181	23
11	0.0966	0.0158	16	0.0808	84	0.0966	0.0333	34	0.0080	8	0.0308	32	0.0015	2	0.0132	24
12	0.0437	0.0071	16	0.0366	84	0.0437	0.0235	54	0.0065	15	0.0101	23	+0.0090	-21	0.0126	29
13	0.0445	0.0067	15	0.0378	85	0.0445	0.0269	60	0.0054	12	0.0066	15	0.0163	37	+0.0107	-24
14	0.0368	0.0099	27	0.0269	73	0.0368	0.0042	11	0.0019	5	0.0126	34	0.0426	116	+0.0132	-66
15	0.0259	0.0054	21	0.0205	79	0.0259	0.0150	58	0.0047	18	0.0135	52	+0.0052	-20	+0.0021	-8
16	0.1080	0.0536	50	0.0544	50	0.1080	0.0405	38	0.0392	36	0.0250	23	0.0049	5	+0.0016	-2

APPENDIX 3

BOTTOM ASH ANALYSES

Run No.	C %	S %	Na %	Ca %	V mg/kg	Ni mg/kg	Part. size 50% less than mm
4	0.60	5.27	0.64	23.2	520	120	
5	0.83	11.12	0.22	22.6	300	50	
6	0.91	8.35	0.25	26.5	430	110	
7a	1.10	11.44	0.20	25.2	740	170	
8a	0.78	12.27	0.30	20.6	600	100	
7b	2.38	9.12	0.20	23.7	840	220	0.24
8b	1.00	11.65	0.28	25.1	750	300	0.35
9	0.52	11.70	0.35	27.0	500	500	0.44
10	1.30	11.59	0.20	23.2	910	170	
11	1.91	10.53	0.32	22.0	600	100	0.33
12	1.03	8.77	0.28	17.5	600	100	0.30
13	0.89	9.95	0.48	19.0	700	300	0.39
14	0.77	5.36	0.47	11.3	400	80	0.23
15	0.87	7.39	0.25	20.5	910	220	
16	1.34	10.14	0.45	22.0	600	300	0.34

APPENDIX 4

Boiler ash analyses

6.6.5 **Vanadium balance**

As can be seen from Table 14, most of the vanadium entering the reactor is discharged in the boiler ash and fly ash, i.e. in the ash with the smallest particle sizing. About 5-15% of the vanadium is drained with the reactor bottom ash. From this distribution, we infer that most of the vanadium compounds entering the reactor were vaporised and condensed upon the ash particles at lower temperature. Fly ash particles leaving the boiler, having the lowest temperature, collected the bulk (40-80%) of the condensed vanadium compounds.

The variation of vanadium content in the fly ash (range 23-83%) may be the result of ash accumulation in the boiler section, and also due to errors in sampling and analysis.

Referring back to Table 10 (page 44), the vanadium in the particulates measured in the baghouse exhaust duct amount to less than 1% of the total vanadium input to the reactor. The high values for test runs 2 and 3 result from the baghouse being by-passed, and particulates that would normally be trapped by the baghouse were recorded by the particulate sampler.

No attempt was made to collect and/or measure gaseous or vaporised vanadium compounds in the stack gases. These are expected to be minor amounts.

Table 14. Vanadium balance

Run No.	INPUT Total	V in coke		OUTPUT Total	V in bottom ash		V in boiler ash		V in baghouse ash		V variation in bed		Accumulation + error	
	g/s	g/s	% of input	g/s	g/s	% of input	g/s	% of input	g/s	% of input	g/s	% of input	g/s	% of input
0	0.0587	0.0587	100	0.0587	-	-	0.0085	14	0.0324	55	0.0000	0	0.0178	31
1	0.0661	0.0661	100	0.0661	-	-	0.0053	8	0.0252	30	0.0025	4	0.0331	50
2	0.0621	0.0621	100	0.0621	-	-	0.0041	7	0.0425	68	0.0004	1	0.0151	24
3	0.0737	0.0737	100	0.0737	-	-	0.0024	3	0.0459	62	-0.0016	-2	0.0270	37
4	0.0695	0.0695	100	0.0695	0.0048	7	0.0021	3	0.0578	83	-0.000.	0	0.0051	7
5	0.0730	0.0730	100	0.0730	0.0032	4	0.0012	2	0.0573	79	0.0010	1	0.0103	14
6	0.0753	0.0753	100	0.0753	0.0063	8	0.0019	3	0.0529	70	0.0018	2	0.0124	17
7a	0.0692	0.0692	100	0.0692	0.0080	12	0.0035	5	0.0539	78	-0.0008	-1	0.0046	6
8a	0.0648	0.0648	100	0.0648	0.0071	11	0.0018	3	0.0377	58	0.0038	6	0.0144	22
7b	0.0647	0.0647	100	0.0647	0.0123	19	0.0049	8	0.0221	34	0.0060	9	0.0194	30
8b	0.0677	0.0677	100	0.0677	0.0079	12	0.0157	23	0.0154	23	0.0004	1	0.0283	41
9	0.0700	0.0700	100	0.0700	0.0049	7	0.0119	17	0.0540	77	0.0004	1	-0.0012	-2
10	0.0722	0.0722	100	0.0722	0.0094	13	0.0028	4	0.0433	60	-0.0029	-4	0.0196	27
11	0.0780	0.0780	100	0.0780	0.0062	8	0.0083	11	0.0300	38	-0.0006	-1	0.0341	44
12	0.0466	0.0466	100	0.0466	0.0050	11	0.0036	8	0.0182	39	0.0014	3	0.0184	39
13	0.0432	0.0432	100	0.0432	0.0049	11	0.0042	10	0.0143	33	0.0021	5	0.0177	39
14	0.0385	0.0385	100	0.0385	0.0004	1	0.0024	6	0.0148	38	0.0101	26	0.0108	29
15	0.0350	0.0350	100	0.0350	0.0055	16	0.0030	9	0.0240	69	0.0002	1	0.0023	5
16	0.0656	0.0656	100	0.0656	0.0054	8	0.0136	21	0.0343	52	0.0001	0	0.0122	19

6.6.6 **Nickel balance**

The nickel distribution follows the distribution of vanadium quite closely, i.e. a concentration of nickel occurs on the smaller fly ash particles with higher surface area and lower temperature. This is illustrated in Table 15 and in Appendices 3, 4 and 5. The concentrations of nickel on the particulates escaping the baghouse are of a similar magnitude to the concentrations on the baghouse ash; this is to be expected.

Table 15. Nickel balance

Run No.	INPUT Total	Ni in coke		OUTPUT Total	Ni in bottom ash		Ni in boiler ash		Ni in baghouse ash		Ni variation in bed		Accumulation + error	
	g/s	g/s	% of input	g/s	g/s	% of input	g/s	% of input	g/s	% of input	g/s	% of input	g/s	% of input
0	0.0175	0.0175	100	0.0175	-	-	0.0030	17	0.0115	66	-0.0004	-2	0.0034	19
1	0.0197	0.0197	100	0.0197	-	-	0.0017	9	0.0056	28	0.0010	5	0.0114	58
2	0.0185	0.0185	100	0.0185	-	-	0.0015	8	0.0150	81	-0.0008	-4	0.0028	15
3	0.0195	0.0195	100	0.0195	-	-	0.0010	5	0.0132	68	-0.0004	-2	0.0057	29
4	0.0202	0.0202	100	0.0202	0.0011	5	0.0011	5	0.0240	119	-0.0057	28	-0.0003	-1
5	0.0219	0.0219	100	0.0219	0.0005	2	0.0010	5	0.0128	58	0.0005	2	0.0071	33
6	0.0208	0.0208	100	0.0208	0.0016	8	0.0009	4	0.0162	78	0.0011	5	0.0010	5
7a	0.0187	0.0187	100	0.0187	0.0018	10	0.0024	13	0.0139	74	-0.0001	-1	0.0007	4
8a	0.0232	0.0232	100	0.0232	0.0011	5	0.0008	3	0.0147	63	0.0017	7	0.0049	22
7b	0.0186	0.0186	100	0.0186	0.0032	17	0.0010	5	0.0062	33	0.0016	9	0.0068	36
8b	0.0249	0.0249	100	0.0249	0.0029	12	0.0045	18	0.0064	26	-0.0001	0	0.0112	44
9	0.0255	0.0255	100	0.0255	0.0079	31	0.0042	16	0.0192	75	-0.0021	-8	-0.0037	-14
10	0.0275	0.0275	100	0.0275	0.0027	10	0.0011	4	0.0245	89	-0.0040	-15	0.0032	12
11	0.0288	0.0288	100	0.0288	0.0010	3	0.0025	9	0.0120	42	0.0084	29	0.0049	17
12	0.0175	0.0175	100	0.0175	0.0008	5	0.0018	10	0.0076	43	-0.0016	-9	0.0089	51
13	0.0166	0.0166	100	0.0166	0.0017	10	0.0017	10	0.0056	34	0.000	0	0.0076	46
14	0.0143	0.0143	100	0.0143	0.0001	1	0.0010	7	0.0061	43	0.0043	30	0.0028	19
15	0.0131	0.0131	100	0.0131	0.0013	10	0.0012	9	0.0122	93	0.0010	8	-0.0026	-20
16	0.0248	0.0248	100	0.0248	0.0027	11	0.0048	19	0.0137	55	0.000	0	0.0036	15

7. BOILER DESIGN CONSIDERATIONS

The results from the test program have been evaluated with a view to considering their effect upon the design of a commercial boiler. The following comments discuss the test results as they relate to commercial boiler design.

The design of an Ahlstrom PYROFLOW CFB boiler requires a number of input parameters to be determined from the proposed fuel and limestone, before an accurate design with full commercial warranties can be made. Clearly, if the fuel is a standard fuel, such as coal or oil, sufficient design information exists for an accurate design to be developed without the aid of test burns. The Syncrude coke falls into the category of requiring a test burn, and the test work described in this test report is of sufficient depth to enable full commercial designs of PYROFLOW boilers to be offered to potential customers.

The design of the PYROFLOW boiler requires the following detailed information to be available for the fuel and limestone:

a. combustor operating temperature
b. primary/secondary air ratio
c. ash split - bottom ash to fly ash
d. Ca:S molar ratio
e. SO_2 emission control requirement
f. NOx emission control requirement
g. combustion efficiency
h. fuel characteristics (burning rate)
i. slagging/fouling
j. corrosion/erosion
k. heat transfer characteristics

Clearly, some of the above information is proprietary to the PYROFLOW process and will not be detailed in this report.

a. Combustor operating temperature

The Syncrude coke is a low volatile fuel (5.6%) and experience has shown that in order to obtain stable combustion characteristics the operating temperature should be slightly higher than would normally be used for a conventional coal in a CFB boiler (about 850°C). From the test work, the optimum combustor temperature was found to be about 900°C. This temperature was selected to satisfy both the combustion, sulphur retention and NOx emission requirements.

b. Primary/secondary air ratio

The adjustment of this ratio controls the degree of combustion that occurs towards the bottom of the combustor. It also stages the combustion process, which enables good control of NOx emissions to be achieved. The test work indicated that a

primary/secondary air ratio of about 1.5/1 (60% primary air/40% secondary air) provides a good balance between good combustion control and low NOx emissions. Clearly, because of the low volatile content of the fuel, more primary air is required than for other more reactive fuels, and NOx control is compromised.

c. Ash split

The division between bottom ash and fly ash is required in order to design the bottom ash handling and cooling system, together with the sizing of the baghouse or electrostatic precipitator. The test work indicated that, with limestone addition for sulphur capture, the ash split is about 50/50.

d. Ca:S molar ratio

The Ca:S molar ratio determines the limestone feedrate for the required sulphur retention level. This influences the limestone storage and handling equipment design. The Fort McMurray limestone proved to be fairly reactive, and a Ca:S molar ratio of about 1.7:1 was found to be required to meet the target SO_2 retention level of 90%. Thus, the limestone equipment needs will not be excessive or onerous.

e. SO_2 emission requirements

The 90% sulphur capture requirement was found to be achieved quite comfortably with the Ahlstrom PYROFLOW boiler, and high utilisation of the limestone was achieved. Ca:S molar ratio ranged from 1.03 to 2.26, with a mean of about 1.65. In a commercial boiler, with prolonged operation, it may be possible to lower the Ca:S molar ratio somewhat.

f. NOx emission requirements

The test work showed that at optimum temperature (about 900°C) low NOx emission levels can be achieved simply by controlling the primary/secondary air ratio. NOx emission levels of lower than 100 ppm (0.064 mg/kJ) were demonstrated.

g. Combustion efficiency

The combustion efficiency was high for such a low volatile fuel, indicating that Syncrude coke can be burned efficiently in a CFB boiler. With such an unreactive fuel, the advantages of CFB technology are clear. Combustion efficiencies during the test runs ranged from 96.2-99.4%, with the highest values occurring in the last 9 tests (average 99%).

h. Fuel characteristics

The fuel characteristics indicated that although the fuel contains low volatiles, and could be expected to be unreactive, it burns well in the lower portion of the combustor, rather than higher up, as can be exhibited by other more unreactive fuels such as, for example, anthracite culm. For unreactive fuels, such as anthracite culm, this higher burning level in the combustor can lead to excessive combustion carry-over into the cyclone if the unit is improperly designed. The tests indicated that, although there is a slight amount of combustion carry-over, this is not expected to be a significant concern for Syncrude coke in a CFB boiler.

i. Slagging/fouling

The test work indicated that, with limestone addition, the ash is very alkaline and easily agglomerated. This may be the result of the high alkali content of the limestone and/or because of the high vanadium and nickel content of the ash. This ease of agglomeration and stickiness of the ash indicates that the design of the superheater surfaces and other large convection surfaces will require slightly greater tube spacing than normal to prevent ash build-up on the surfaces. The use of sootblowers will be required; the sonic type could prove to be the most suitable for these ash deposits. Air sootblowers could also be used.

j. Corrosion/erosion

The tests were not of sufficient duration to determine the corrosion/erosion potential of the ash. However, examination of the ash did indicate that this concern should be minimal.

The ash fusion temperature (1270-1290^{o}C) and the combustion temperature (900^{o}C) should ensure that no ash slagging will occur in the combustor, and that corrosion from fused ash (as has been experienced in other tar sands boilers) will not be a significant problem.

h. Heat transfer characteristics

The heat transfer coefficient is a function of many parameters operating within the CFB. Testwork has shown that the parameters having the major impact upon overall heat transfer coefficient within the combustor are: 1) operating temperature; 2) superficial fluidising velocity; 3) solids circulation rate; and 4) mean particle size of combustor solids.

The range of heat transfer coefficients (k) determined from the test runs was 70-170 W/m^2/^{o}K for the overall system heat transfer.

APPENDICES

APPENDIX 1

Syncrude coke analyses

APPENDIX 1

SYNCRUDE TEST

COKE ANALYSES

Run No.	Dry solids (d.s.)%	Ash in d.s %	C in d.s %	S in d.s %	Ca in d.s mg/kg	Na in d.s mg/kg	V in d.s mg/kg	Ni in d.s mg/kg	Part. size 50% less than mm
0	95.6	8.30	76.9	6.72	400	500	1980	590	0.16
1	95.7	9.44	77.7	6.64	900	1000	1980	590	0.18
2	96.2	7.73	77.7	6.78	1300	500	2010	600	0.22
3	96.9	7.89	78.6	6.56	1100	700	2090	580	0.15
4	97.1	6.97	76.7	6.80	2000	200	2070	600	0.21
5	97.4	6.84	77.9	6.64	2300	300	2030	610	0.17
6	96.9	7.43	77.5	6.64	2400	300	2140	590	
7a	97.5	6.85	78.6	6.72	1000	200	2150	580	0.18
8a	97.1	7.21	77.7	6.66	1400	500	1900	680	0.23
7b	96.3	6.65	79.8	6.82	1400	200	2060	600	
8b	96.8	7.10	79.3	6.80	600	500	2040	750	
9	96.9	6.76	80.3	6.80	1000	200	2000	730	
10	96.7	7.45	80.2	6.89	1500	300	1920	730	
11	96.3	7.86	79.5	6.77	1300	400	1980	730	
12	96.6	8.38	79.9	6.88	2300	300	1970	740	
13	96.5	8.65	79.1	6.83	1900	300	1930	740	
14	97.3	7.22	79.77	6.72	2000	500	1940	720	
15	96.5	8.25	79.4	6.71	1500	300	1950	730	
16	96.7	7.71	79.6	6.89	1500	1600	1960	740	

PARTICLE SIZE DISTRIBUTION (single composite sample)

Sieve size (mm)	Weight % passing sieve
2.0	99.5
1.0	99.2
0.5	98.2
0.35	96.5
0.25	66.8
0.18	29.4
0.125	6.7
0.063	0.1

APPENDIX 2

Limestone analyses

APPENDIX 2

LIMESTONE ANALYSES

1. OVERALL SAMPLES

Sample	Ca in d.s. %	Mg in d.s. %	CO_3 in d.s. %	C in d.s. %	S in d.s. %	Na in d.s. %	K in d.s. %	Cl in d.s. %	Moisture %	Loss of ign. %	Part. size 50% less than mm
A	31.4	0.60	48.3	9.13	0.18	0.19	0.87	0.27	0.07	34.47	0.35
B	31.3	0.57	54.8	9.21	0.16	0.25	0.90	0.33	0.06	34.59	0.28
C	31.0	0.61	54.7	9.19	0.16	0.23	0.89	0.26	0.06	34.59	0.31

2. SAMPLES OF EACH TEST RUN

Run No.	Ca in d.s %	Part. size 50% less than mm	Na in d.s. %	K in d.s. %	V in d.s. mg/kg	Ni in d.s. mg/kg
4	29.0	0.25	0.39	1.0	0	0
5	29.8		0.35	1.0	0	0
6	31.0	0.66	0.39	0.70	0	0
7a	28.0	0.36	0.35	0.70	0	0
8a	28.8	0.42	0.38	0.80	0	0
7b	29.8	0.31	0.36	0.60	0	0
8b	31.0	0.42	0.44	0.86	0	0
9	29.0		0.43	1.0	0	0
10	29.0	0.32	0.35	0.95	0	0
11	29.0	0.35	0.40	0.93	0	0
12	27.5	0.50	0.37	0.89	0	0
13	30.0	0.30	0.42	0.80	0	0
14	31.6	0.35	0.42	0.76	0	0
15	31.0		0.41	1.3	0	0
16	30.1	0.34	0.37	0.78	0	0

d.s. = dry solids

3. PARTICLE SIZE DISTRIBUTION (single composite sample)

Sieve size (mm)	Weight % passing sieve
2.0	97.9
1.0	84.2
0.5	60.8
0.35	54.6
0.25	41.3
0.18	34.9
0.09	24.9
0.044	19.6

APPENDIX 3

Bottom ash analyses

APPENDIX 4

SYNCRUDE TEST
BOILER ASH ANALYSES

Run No.	C %	S %	Na %	Ca %	V %	Ni %	Part. size 50% less than μm
0	22.53	2.55	0.71	1.0	0.34	0.12	45
1	26.06	3.14	0.72	1.0	0.41	0.13	30
2	19.33	2.39	0.92	1.3	0.41	0.15	34
3	10.85	1.41	0.80	1.2	0.40	0.16	45
4	4.54	8.06	0.70	17.0	0.19	0.10	33
5	6.38	10.43	0.48	20.8	0.23	0.19	30
6	8.78	8.77	0.49	21.1	0.27	0.13	25
7a	5.80	8.44	0.48	22.9	0.25	0.17	18
8a	6.24	9.47	0.26	19.8	0.22	0.06	23
7b	6.40	8.56	0.44	28.9	0.27	0.10	15
8b	4.38	7.50	0.41	19.7	0.49	0.14	11
9	2.97	8.33	0.42	21.6	0.34	0.12	15
10	3.80	7.85	0.48	22.2	0.25	0.10	19
11	2.98	7.37	0.32	19.2	0.33	0.10	18
12	3.32	8.11	0.30	20.3	0.20	0.10	14
13	4.85	7.62	0.54	19.0	0.42	0.17	15
14	5.80	7.40	0.32	19.5	0.40	0.17	15
15	5.22	7.36	0.43	17.9	0.28	0.11	13
16	3.86	8.62	0.80	19.0	0.31	0.11	15

PARTICLE SIZE DISTRIBUTION

Data from a Malvern Instruments EASY Particle Sizer

Particle size (microns)	Weight % less than particle size			
	Run 10	Run 12	Run 15	Run 16
118.4	100.0	100.0	100.0	100.0
102.2	99.6	100.0	99.8	99.7
88.2	98.6	100.0	99.4	98.9
76.0	97.2	100.0	98.8	97.9
65.6	95.8	100.0	98.3	96.8
56.6	94.6	99.2	97.9	96.0
48.8	92.4	97.2	96.4	94.0
42.1	87.8	94.1	92.7	89.6
36.3	81.6	90.2	87.9	84.0
31.4	74.9	85.4	83.0	78.1
27.1	67.9	80.0	78.0	72.3
23.3	60.7	73.9	72.0	65.8
20.2	53.8	67.8	65.3	58.9
17.3	47.2	61.5	58.9	52.7
15.0	41.5	55.4	53.9	48.3
12.9	35.1	47.6	47.0	42.7
11.1	28.4	39.1	38.3	35.6
9.6	23.5	33.1	32.2	30.3
8.3	20.2	29.4	29.0	27.1
7.2	17.5	26.0	26.4	24.7
6.2	14.9	22.1	23.0	21.9
5.3	12.2	17.9	18.8	18.1
4.6	10.0	15.0	15.7	14.9
4.0	8.6	13.7	14.3	13.0
3.4	7.6	12.8	13.6	11.9
3.0	6.9	12.1	13.2	11.4
2.6	6.3	11.5	12.8	10.9
2.2	5.6	10.2	11.4	9.4
1.9	4 6	7.9	8.5	6.8
1.6	3.4	4.5	4.4	3.6
1.4	2.3	2.1	1.7	1.5
1.2	1.5	1.3	1.1	1.0

Run 10 was maximum load with no baghouse ash recirculation
Run 12 was intermediate load with baghouse ash recirculation
Run 15 was minimum load with baghouse ash recirculation
Run 16 was long nominal load test with baghouse ash recirculation

APPENDIX 5

Baghouse ash analyses

APPENDIX 5

SYNCRUDE TEST
BAGHOUSE ASH ANALYSES

Run No.	C %	S %	Na %	Ca %	V %	Ni %	Part size 50% less than µm
0	21.28	2.79	0.72	2.1	1.8	0.46	
1	23.66	3.16	0.91	1.5	1.8	0.40	
2	*16.12	2.91	0.83	2.2	1.7	0.60	
3	*11.08	2.24	0.68	2.3	1.7	0.49	
4	3.95	5.86	0.74	17.8	0.53	0.22	
5	5.53	6.37	0.48	20.0	0.94	0.21	
6	8.32	5.96	0.73	18.9	0.98	0.30	
7a	4.76	6.63	0.49	20.9	0.70	0.18	5
8a	6.01	6.96	0.45	17.7	0.77	0.30	6
7b	3.50	6.66	0.66	18.0	0.82	0.23	2
8b	2.00	5.08	0.60	17.0	0.96	0.40	2
9	1.82	6.77	0.69	21.0	0.76	0.27	6
10	3.78	6.22	0.50	18.6	0.76	0.43	7
11	3.10	6.37	0.77	20.1	0.75	0.30	7
12	1.56	6.15	0.53	18.6	0.96	0.40	2
13	1.80	5.62	0.51	18.9	1.1	0.43	3
14	3.00	5.98	0.74	23.0	0.87	0.36	5
15	1.17	6.65	0.50	19.8	0.89	0.45	4
16	2.85	5.90	0.64	22.6	0.88	0.35	4

*measured from particulates, baghouse bypassed

PARTICLE SIZE DISTRIBUTION

Data from a Malvern Instruments EASY Particle Sizer

Particle size (microns)	Weight % less than particle size			
	Run 10	Run 12	Run 15	Run 16
118.4	100.0	100.0	100.0	100.0
102.2	100.0	100.0	100.0	100.0
88.2	100.0	100.0	100.0	100.0
76.0	100.0	100.0	100.0	99.9
65.5	99.3	100.0	100.0	99.7
56.6	98.3	100.0	100.0	99.4
48.8	96.9	100.0	100.0	99.0
42.1	94.9	100.0	100.0	98.3
36.3	92.5	100.0	100.0	97.5
31.4	89.7	100.0	99.9	96.4
27.1	86.6	100.0	99.9	95.1
23.3	83.2	99.9	99.7	93.4
20.2	79.8	99.8	99.4	91.5
17.3	76.3	99.6	98.8	89.2
15.0	73.0	99.2	98.0	86.9
12.9	68.7	97.8	96.2	84.0
11.1	63.9	95.2	93.4	80.5
9.6	59.8	92.0	89.9	76.9
8.3	56.2	88.2	85.6	72.9
7.2	50.4	83.4	80.4	68.7
6.2	48.1	77.4	74.0	63.9
5.3	42.7	70.1	66.4	58.4
4.6	37.6	63.6	59.2	52.6
4.0	33.7	59.0	53.4	47.1
3.4	31.1	56.3	49.2	42.1
3.0	29.5	54.4	45.6	36.7
2.6	28.0	52.4	41.7	31.0
2.2	17.3	32.7	25.1	19.1
1.9	0.6	0.8	1.1	4.3
1.6	0.2	0.2	0.3	1.0
1.4	0	0	0	0
1.2	0	0	0	0

Run 10 was maximum loal with no baghouse ash recirculation
Run 12 was intermediate load with baghouse ash recirculation
Run 15 was minimum load with baghouse ash recirculation
Run 16 was nominal load with baghouse ash recirculation

APPENDIX 6

Bed material analyses

APPENDIX 6

SYNCRUDE TEST
BED MATERIAL ANALYSES **

Run No.	C %	S %	Na %	Ca %	V mg/kg	Ni mg/kg	Part. size 50% less than mm
0	0.51	0.44	0.62	0.88	710	350	
	0.49	0.60	0.56	0.93	750	320	
1	0.55	0.60	0.66	0.90	680	610	
	0.49	0.61	0.65	0.94	720	600	
2	0.54	0.78	0.63	1.3	520	300	
	0.85	0.88	0.56	0.85	670	230	
3	0.44	0.65	0.51	1.5	740	370	
	0.47	0.60	0.47	1.4	620	350	
4	0.52	7.23	0.43	16.7	480	610	
	0.57	10.04	0.09	22.3	420	70	
5	0.61	9.98	0.05	24.7	390	70	
	0.87	11.68	0.06	26.1	470	110	
6	1.09	6.10	0.10	21.6	460	370	
	0.79	9.36	0.09	25.3	520	390	
7a	0.88	12.05	0.08	24.0	700	150	
	0.92	10.51	0.07	23.6	670	150	
8a	0.70	11.77	0.11	23.4	520	130	
	0.57	12.37	0.06	22.4	670	220	
7b	0.96	9.34	0.08	30.4	420	120	
	1.10	9.51	0.12	29.2	940	260	
8b	0.83	10.81	0.08	25.4	520	130	
	0.70	12.40	0.05	23.7	530	120	
9	0.78	11.65	0.05	28.2	520	210	
	0.38	13.38	0.04	25.6	560	60	
10	0.33	11.53	0.06	25.1	410	170	0.44
	0.26	11.64	0.05	25.3	350	170	
11	0.35	12.26	0.05	26.8	350	740	
	0.38	11.71	0.05	26.7	310	890	
12	0.71	9.02	0.38	26.0	580	200	
	0.71	9.04	0.35	26.0	670	130	0.26
13	0.65	9.24	0.30	23.4	900	100	
	0.88	9.33	0.38	30.0	940	90	
14	0.17	1.46	0.80	8.0	140	70	
	0.98	5.97	0.81	10.4	470	210	0.20
15	1.00	6.95	0.56	17.6	530	180	
	0.73	7.83	0.60	18.0	600	250	
16	1.01	11.32	0.26	26.0	420	110	
	0.85	10.72	0.33	30.0	390	90	0.48

** The first (upper) numbers for each run are the concentrations at the start of the steady-state period for each test run. The lower numbers are the values at the end of each run.

PARTICLE SIZE DISTRIBUTION

Sieve size (mm)	Weight % passing sieve Run 10	Run 12	Run 14	Run 16
2.0	97.6	97.7	99.2	97.7
1.0	85.1	88.5	94.9	83.1
0.5	56.4	67.7	82.8	50.5
0.355	40.4	54.7	76.5	38.4
0.25	25.5	39.3	54.2	18.5
0.18	13.8	24.3	35.2	10.3
0.09	1.9	4.6	5.8	2.1
0.045	0.1	0.5	0.5	0.2

Run 10 was maximum load with no baghouse ash recirculation
Run 12 was intermediate load with baghouse ash recirculation
Run 15 was minimum load with baghouse ash recirculation
Run 16 was nominal load with baghouse ash recirculation

APPENDIX 7

Recirculation material analyses

APPENDIX 7

SYNCRUDE TEST
RECIRCULATION MATERIAL ANALYSES

Run No.	C %	S %	Na %	Ca %	Ni %	V %
0	0.67	0.26	0.80	0.97	0.038	0.085
1	0.84	0.33	0.82	1.0	0.028	0.066
2	1.32	0.42	0.61	1.2	0.028	0.065
3	0.68	0.34	0.72	1.0	0.030	0.079
4	0.68	6.74	0.29	16.8	0.014	0.062
5	1.08	10.53	0.16	24.0	0.011	0.057
6	1.69	8.55	0.26	25.0	0.041	0.071
7a	1.78	12.18	0.15	25.1	0.038	0.069
8a	0.92	11.88	0.10	24.7	0.045	0.076
7b	1.15	10.24	0.16	20.9	0.032	0.065
8b	0.87	10.91	0.12	27.1	0.025	0.062
9	0.48	12.24	0.10	23.5	0.021	0.059
10	0.35	11.14	0.12	25.0	0.023	0.065
11	0.27	11.49	0.11	29.9	0.020	0.060
12	1.03	8.48	0.31	16.5	0.029	0.075
13	1.15	8.62	0.22	15.7	0.040	0.081
14	1.23	6.01	0.49	11.2	0.032	0.080
15	1.20	7.17	0.62	13.5	0.035	0.072
16	0.77	11.77	0.48	27.8	0.027	0.076

APPENDIX 8

Particulate analyses
(Solids passing baghouse)

APPENDIX 8

SYNCRUDE TEST
PARTICULATE ANALYSES

Run No.	Ca %	Na %	V %	Ni %
0	2.4	0.91	2.3	0.85
1	1.9	0.76	1.9	1.10
2	2.2	0.82	1.7	0.60
3	2.3	0.88	1.7	0.49
4	9.6	0.52	0.36	0.07
5	12.1	0.48	0.52	0.16
8a	10.5	0.42	0.79	0.26
7b	8.1	0.30	0.80	0.30
9	8.6	0.39	0.74	0.26
10	10.0	0.64	0.60	0.18
11	9.6	0.29	0.18	0.06
12	7.3	0.36	0.30	0.10
13	6.9	0.40	0.59	0.27
14	8.8	0.27	0.31	0.09
16	13.1	0.61	0.73	0.20